

Reverse Osmosis

Industrial Applications and Processes



Jane Kucera

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Reverse Osmosis

Design, Processes, and
Applications for Engineers

Jane Kucera



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For my dad; he'll always be O.K.

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Preface

The use of reverse osmosis (RO) technology has grown rapidly through the 1990's and early 2000's. The ability of RO to replace or augment conventional ion exchange saves end users the need to store, handle, and dispose of large amounts of acid and caustic, making RO a "greener" technology. Additionally, costs for membranes have declined significantly since the introduction of interfacial composite membranes in the 1980's, adding to the attractiveness of RO. Membrane productivity and salt rejection have both increased, reducing the size of RO systems and minimizing the amount of post treatment necessary to achieve desired product quality.

Unfortunately, knowledge about RO has not kept pace with the growth in technology and use. Operators and others familiar with ion exchange technology are often faced with an RO system with little or no training. This has resulted in poor performance of RO systems and perpetuation of misconceptions about RO.

Much of the current literature about RO includes lengthy discussions or focuses on a niche application that makes it difficult to find an answer to a practical question or problems associated with more common applications. Hence, my objective in writing this book is to bring clear, concise, and practical information about RO to end users, applications engineers, and consultants. In essence, the book is a reference bringing together knowledge from other references as well as that gained through personal experience.

The book focuses on brackish water industrial RO, but many principles apply to seawater RO and process water as well.

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Acknowledgements

My enthusiasm for reverse osmosis (RO) began while working with my thesis advisor at UCLA, Professor Julius “Bud” Glater, a pioneer who worked at UCLA with Sidney Loeb in the early days of commercializing RO. Professor Glater was kind enough to extend a Research Assistantship to me, when my first choice was not available. That was fortunate for me, as membrane technology is a growing field with great future potential. Professor Glater’s guidance and support were invaluable to me as a graduate student and has continued to be throughout my career.

My knowledge grew at Bend Research, Inc. under Harry Lonsdale, another membrane pioneer who was involved in the theoretical and practical side of membranes since the early 1960’s at Gulf General Atomic (predecessor of Fluid Systems, now Koch Membrane Systems), Alza, and later Bend Research, which he co-founded with Richard Baker. At Bend Research, I had the opportunity to develop novel membranes and membrane-based separation processes, including leading several membrane-based projects for water recovery and reuse aboard the International Space Station.

My desire to write this book was fostered by Loraine Huchler, president of Mar-Tech Systems, which she founded in the mid 1990’s, and author of the book series, *Operating Practices for Industrial Water Management*. Loraine has provided both technical and moral support.

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1

FUNDAMENTALS

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Introduction and History of Development

1.1 Introduction

Reverse Osmosis (RO) is a membrane-based demineralization technique used to separate dissolved solids, such as ions, from solution (most applications involve water-based solutions, which is the focus of this work). Membranes in general act as perm-selective barriers, barriers that allow some species (such as water) to selectively permeate through them while selectively retaining other dissolved species (such as ions). Figure 1.1 shows how RO perm-selectivity compares to many other membrane-based and conventional filtration techniques. As shown in the figure, RO offers the finest filtration currently available, rejecting most dissolved solids as well as suspended solids. (Note that although RO membranes will remove suspended solids, these solids, if present in RO feed water, will collect on the membrane surface and foul the membrane. See Chapters 3.7 and 7 for more discussion on membrane fouling).

1.1.1 Uses of Reverse Osmosis

Reverse osmosis can be used to either purify water or to concentrate and recover dissolved solids in the feed water (known as “dewatering”). The most common application of RO is to replace ion exchange, including sodium softening, to purify water for use as boiler make-up to low- to medium-pressure boilers, as the product quality from an RO can directly meet the boiler make-up requirements for these pressures. For higher-pressure boilers and steam generators, RO is used in conjunction with ion exchange, usually as a pretreatment to a two-bed or mixed-bed ion exchange system. The use of RO prior to ion exchange can significantly reduce the frequency of resin regenerations, and hence, drastically reduce the amount of acid, caustic, and regeneration waste that must be handled and stored. In some cases, a secondary RO unit can be used in place of ion exchange to further purify product water from an RO unit (see Chapter 5.3). Effluent from

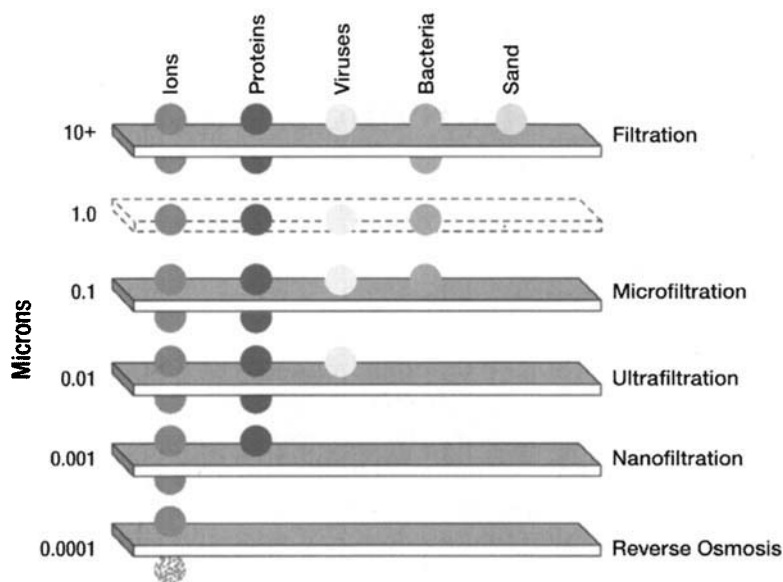


Figure 1.1 “Filtration Spectrum” comparing the rejection capabilities of reverse osmosis with other membrane technologies and with the separation afforded by conventional filtration.

the second RO may be used directly or is sometimes polished with mixed-bed ion exchange or continuous electrodeionization to achieve even higher product water purity (see Chapter 16.3).

Other common applications of RO include:

1. Desalination of seawater and brackish water for potable use. This is very common in coastal areas and the Middle East where supply of fresh water is scarce.
2. Generation of ultrapure water for the microelectronics industry.
3. Generation of high-purity water for pharmaceuticals.
4. Generation of process water for beverages (fruit juices, bottled water, beer).
5. Processing of dairy products.
6. Concentration of corn sweeteners.
7. Waste treatment for the recovery of process materials such as metals for the metal finishing industries, and dyes used in the manufacture of textiles.
8. Water reclamation of municipal and industrial wastewaters.

1.1.2 History of Reverse Osmosis Development

One of the earliest recorded documentation of semipermeable membranes was in 1748, when Abbe Nollet observed the phenomenon of osmosis.¹ Others, including Pfeffer and Traube studied osmotic phenomena using ceramic membranes in the 1850's. However, current technology dates back to the 1940's when Dr. Gerald Hassler at the University of California at Los Angeles (UCLA) began investigation of osmotic properties of cellophane in 1948.² He proposed an "air film" bounded by two cellophane membranes.³ Hassler assumed that osmosis takes place via evaporation at one membrane surface followed by passage through the air gap as a vapor, with condensation on the opposing membrane surface. Today, we know that osmosis does not involve evaporation, but most likely involves solution and diffusion of the solute in the membrane (see Chapter 4).

Figure 1.2 shows a time line with important events in the development of RO technology. Highlights are discussed below.

In 1959, C.E. Reid and E.J. Breton at University of Florida, demonstrated the desalination capabilities of cellulose acetate film.⁴ They evaluated candidate semipermeable membranes in a trial-and-error approach, focusing on polymer films containing hydrophilic groups. Materials tested included cellophane, rubber hydrochloride, polystyrene, and cellulose acetate. Many of these materials exhibited no permeate flow, under pressures as high as 800 psi, and had chloride rejections of less than 35%. Cellulose acetate (specifically the DuPont 88 CA-43), however, exhibited chloride rejections of greater than 96%, even at pressures as low as 400 psi. Fluxes ranged from about 2 gallons per square foot-day (gfd) for a 22-micron thick cellulose acetate film to greater than 14 gfd for a 3.7-micron thick film when tested at 600 psi on a 0.1M sodium chloride solution. Reid and Breton's conclusions were that cellulose acetate showed requisite semipermeability properties for practical application, but that improvements in flux and durability were required for commercial viability.

A decade after Dr. Hassler's efforts, Sidney Loeb and Srinivasa Sourirajan at UCLA attempted an approach to osmosis and reverse osmosis that differed from that of Dr. Hassler. Their approach consisted of pressurizing a solution directly against a flat, plastic film.³ Their work led to the development of the first asymmetric cellulose acetate membrane in 1960 (see Chapter 4.2.1).² This membrane made RO a commercial viability due to the significantly

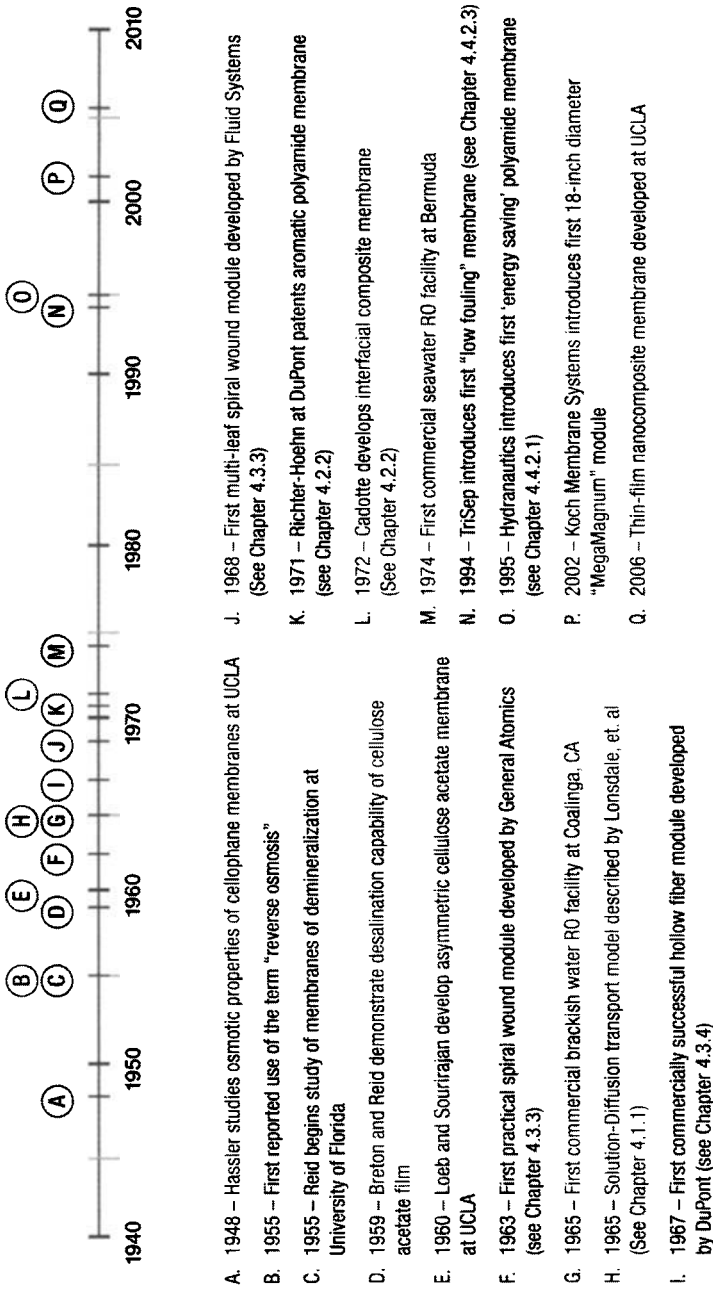


Figure 1.2 Historical time line in the development of reverse osmosis.

improved flux, which was 10 times that of other known membrane materials at the time (such as Reid and Breton's membranes).⁵ These membranes were first cast by hand as flat sheets. Continued development in this area led to casting of tubular membranes. Figure 1.3 is a schematic of the tubular casting equipment used by Loeb and Sourirajan. Figure 1.4 shows the capped, in-floor immersion well that was used by Loeb and students and is still located in Boelter Hall at UCLA.

Following the lead of Loeb and Sourirajan, researchers in the 1960's and early 1970's made rapid progress in the development of commercially-viable RO membranes. Harry Lonsdale, U. Merten, and Robert Riley formulated the "solution-diffusion" model of mass transport through RO membranes (see Chapter 4.1).⁶ Although most membranes at the time were cellulose acetate, this model

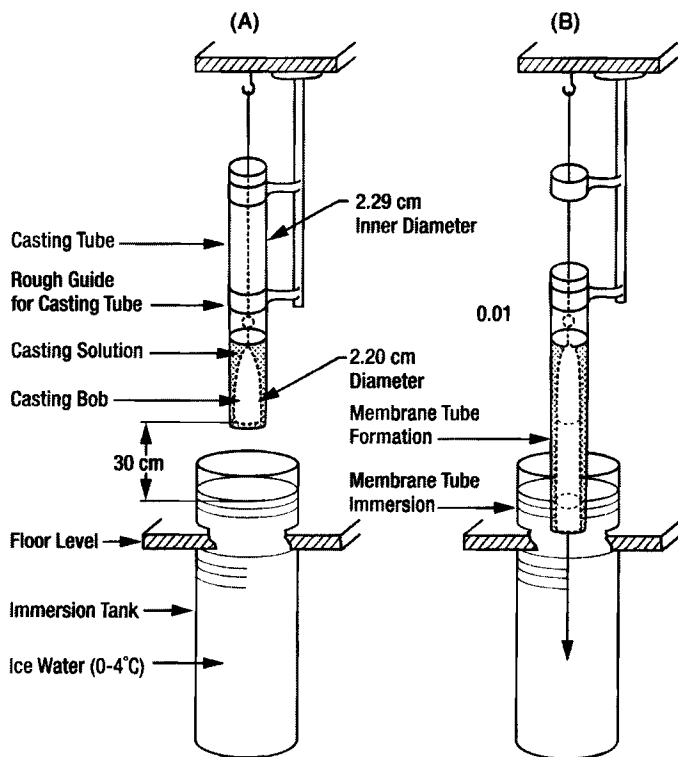


Figure 1.3 Schematic on tubular casting equipment used by Loeb. *Courtesy of Julius Glater, UCLA.*



Figure 1.4 Capped, in-floor immersion tank located at Boelter Hall that was used by Loeb and Sourirajan to cast tubular cellulose acetate membranes at UCLA, as viewed in 2008.

represented empirical data very well, even with respect to present-day polyamide membranes.⁷ Understanding transport mechanisms was important to the development of membranes that exhibit improved performance (flux and rejection).

In 1971, E. I. Du Pont De Nemours & Company, Inc. (DuPont) patented a linear aromatic polyamide with pendant sulfonic acid groups, which they commercialized as the Permasep™ B-9 and B-10 membranes (Permasep is a registered trademark of DuPont Company, Inc. Wilmington, DE). These membranes exhibited higher water flux at slightly lower operating pressures than cellulose acetate membranes. The membranes were cast as unique hollow fine fibers rather than in flat sheets or a tubes (see Chapter 4.3.4).

Cellulose acetate and linear aromatic polyamide membranes were the industry standard until 1972, when John Cadotte, then at North Star Research, prepared the first interfacial composite polyamide membrane.⁸ This new membrane exhibited both higher throughput and rejection of solutes at lower operating pressure than the here-to-date cellulose acetate and linear aromatic polyamide membranes. Later, Cadotte developed a fully aromatic interfacial composite membrane based on the reaction of phenylene diamine and trimesoyl chloride. This membrane became the new industry standard and is known today as FT30, and it is the basis for the majority

of Dow Water and Process Solutions' FilmTec™ membranes (e.g., BW30, which means "Brackish Water membrane," FT30 chemistry"; TW30, which means "Tap Water membrane," FT30 chemistry; and so on) as well as many commercially available membranes from other producers (FilmTec is a trademark of Dow Chemical Company, Midland, Michigan). See Chapter 4.2 for more information about interfacial composite membranes.

Other noteworthy developments in membrane technology include the following:

- 1963: First practical spiral wound module developed at Gulf General Atomics (later known as Fluid Systems®, now owned by Koch Membrane Systems, Wilmington, MA.) This increased the packing density of membrane in a module to reduce the size of the RO system (see Chapter 4.3).
- 1965: The first commercial brackish water RO (BWRO) was on line at the Raintree facility in Coalinga, California. Tubular cellulose acetate membranes developed and prepared at UCLA were used in the facility. Additionally, the hardware for the system was fabricated at UCLA and transported piecemeal to the facility.⁹
- 1967: First commercial hollow-fiber membrane module developed by DuPont. This module configuration further increased the packing density of membrane modules.
- 1968: First multi-leaf spiral wound membrane module developed by Don Bray and others at Gulf General Atomic, under US Patent no. 3,417,870, "Reverse Osmosis Purification Apparatus," December, 1968. A multi-leaf spiral configuration improves the flow characteristics of the RO module by minimizing the pressure drop encountered by permeate as it spirals into the central collection tube.
- 1978: FT-30 membrane patented and assigned to FilmTec (now owned by Dow Chemical Company, Midland, MI).

1.1.3 Recent Advances in RO Membrane Technology

Since the 1970's, the membrane industry has focused on developing membranes that exhibit ever greater rejection of solutes while at

Table 1.1 Development of RO membranes for brackish water desalination.

Year	Pressure (psi)	Relative Flux	Rejection (%)	Membrane Material
1970's	435	1	97	Cellulose acetate
1980's	290	1.9	99.0	Cross-linked polyamide composite
1987	220	3.0	99.7	Cross-linked aromatic polyamide composite
1988	145	4.2	99.7	Cross-linked aromatic polyamide composite
1996	110	5.6	99.7	Cross-linked aromatic polyamide composite
1999	75	8.0	99.7	Cross-linked aromatic polyamide composite

the same time exhibiting higher throughput (flux) at lower operating pressure. Table 1.1 shows the growth in RO membrane development with respect to rejection, flux, and operating pressure.¹⁰ Along with advances in membrane performance, membrane costs have also improved. Table 1.2 lists costs of membranes relative to 1980.⁵

In addition to the progress shown in Table 1.1, some membranes now exhibit up to 99.75% rejection (a drop of 17% in salt passage over membranes exhibiting 99.7% rejection). Other advancements

Table 1.2 Membrane cost decline relative to 1980.⁵

Year	Relative Cost
1980	1.00
1985	0.65
1990	0.34
1995	0.19
2000	0.14

in membrane technology include “low pressure” RO membranes that allow for operation at lower water temperatures ($< 50^{\circ}\text{F}$ (10°C)) with reasonably low operating pressure (see Chapter 4.4.2.1). And, “fouling resistant” membranes have been developed that purport to minimize fouling by suspended solids, organics, and microbes (see Chapter 4.4.2.3).

Since the late 1970's, researchers in the US, Japan, Korea, and other locations have been making an effort to develop chlorine-tolerant RO membranes that exhibit high flux and high rejection. Most work, such as that by Riley and Ridgway et.al., focuses on modifications in the preparation of polyamide composite membranes (see Chapter 4.2.2).¹¹ Other work by Freeman (University of Texas at Austin) and others involves the development of chlorine-tolerant membrane materials other than polyamide. To date, no chlorine-resistant polyamide composite membranes are commercially available for large-scale application.

Nanotechnology came to RO membranes on a research and development scale in the mid 2000's, with the creation of thin-film nanocomposite membranes.^{2,12,13} The novel membranes created at UCLA in 2006 by Dr. Eric M.V. Hoek and team include a type of zeolite nanoparticle dispersed within the polyamide thin film. The nanoparticles have pores that are very hydrophilic such that water permeates through the nanoparticle pores with very little applied pressure as compared to the polyamide film, which requires relatively high pressure for water to permeate. Hence, the water permeability through the nanocomposite membranes at the highest nanoparticle loading investigated, is twice that of a conventional polyamide membrane.¹² The rejection exhibited by the nanocomposite membrane was equivalent to that of the conventional polyamide membrane.¹² The controlled structure of the nanocomposite membrane purports to improve key performance characteristics of reverse osmosis membranes by controlling membrane roughness, hydrophilicity, surface charge, and adhesion of bacteria cells.¹⁴ The thin-film nanocomposite membrane (TFN) technology was licensed from UCLA in 2007 by NanoH₂O, Inc. (Los Angeles, CA) for further research and development toward commercialization.¹⁵

Along similar lines, other researchers have been looking into nanocomposite membranes.¹⁶ Researchers at the University of Colorado at Boulder have been developing lyotropic liquid crystals (LLCs) to form what they call nanostructured polymer membranes.¹⁶ The LLCs can form from liquid crystalline phases with regular geometries

which act as conduits for water transport while rejection ions based on size exclusion. In bench-scale tests, nanostructured polymer membranes exhibited a rejections of 95% and 99.3% of sodium chloride and calcium chloride, respectively.¹³ These membranes also exhibited greater resistance to chlorine degradation than commercially-available polyamide composite membranes. As is the case with the nanocomposite membranes, the nanostructured polymer membranes are not yet in commercial production.

1.1.4 Future Advancements

Improvements will be necessary as RO is used to treat the ever greater expanding candidate feed waters, including municipal and industrial wastewater effluents, and other source waters that are less than optimal for conventional RO membranes (e.g., wastewaters containing high concentrations of biological chemical demand (BOD), chemical oxygen demand (COD), TOC, silica, and suspended solids, such as food-processing condensates and cooling tower blowdown). Membranes will need to be developed that are tolerant of chlorine for microbial growth control, and resist to fouling with suspended solids and organics. Other membrane technologies, such as microfiltration and ultrafiltration, are finding fresh application in pre-treating RO systems operating on these challenging water sources.

There is also continuing research into higher-performance (high flux and high rejection) membranes to further reduce the size and cost of RO systems. Nanotechnology shows promise for having a role in the development of these high-performance membranes.

Improvements will be required in the chemistries used to treat RO. These chemistries include antiscalants, which will be needed to address higher concentrations of scale formers such as silica, and membrane cleaners, which will have to address microbes, biofilms, and organics.

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Reverse Osmosis Principles

Reverse osmosis is a demineralization process that relies on a semipermeable membrane to effect the separation of dissolved solids from a liquid. The semipermeable membrane allows liquid and some ions to pass, but retains the bulk of the dissolved solids. Although many liquids (solvents) may be used, the primary application of RO is water-based systems. Hence, all subsequent discussion and examples will be based on the use of water as the liquid solvent.

To understand how RO works, it is first necessary to understand the natural process of osmosis. This chapter covers the fundamentals of osmosis and reverse osmosis.

2.1 Osmosis

Osmosis is a natural process where water flows through a semipermeable membrane from a solution with a low concentration of dissolved solids to a solution with a high concentration of dissolved solids.

Picture a cell divided into 2 compartments by a semipermeable membrane, as shown in Figure 2.1. This membrane allows water and some ions to pass through it, but is impermeable to most dissolved solids. One compartment in the cell has a solution with a high concentration of dissolved solids while the other compartment has a solution with a low concentration of dissolved solids. Osmosis is the natural process where water will flow from the compartment with the low concentration of dissolved solids to the compartment with the high concentration of dissolved solids. Water will continue to flow through the membrane until the concentration is equalized on both sides of the membrane.

At equilibrium, the concentration of dissolved solids is the same in both compartments (Figure 2.2); there is no more net flow from one compartment to the other. However, the compartment that once contained the higher concentration solution now has a higher water level than the other compartment.

The difference in height between the 2 compartments corresponds to the osmotic pressure of the solution that is now at equilibrium.

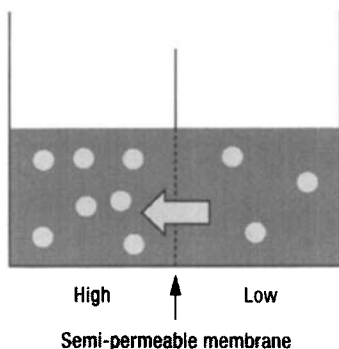


Figure 2.1 Cell divided into 2 compartments separated by a semipermeable membrane. Water moves by osmosis from the low-concentration solution in one compartment through the semipermeable membrane into the high-concentration solution in the other compartment.

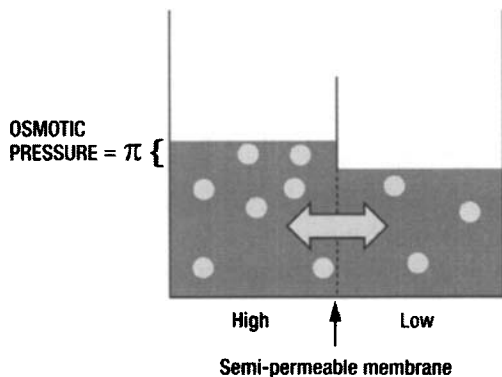


Figure 2.2 Concentration equilibrium. Difference in height corresponds to osmotic pressure of the solution.

Osmotic pressure (typically represented by π (pi)) is a function of the concentration of dissolved solids. It ranges from 0.6 to 1.1 psi for every 100 ppm total dissolved solids (TDS). For example, brackish water at 1,500 ppm TDS would have an osmotic pressure of about 15 psi. Seawater, at 35,000 ppm TDS, would have an osmotic pressure of about 350 psi.

2.2 Reverse Osmosis

Reverse osmosis is the process by which an applied pressure, greater than the osmotic pressure, is exerted on the compartment that

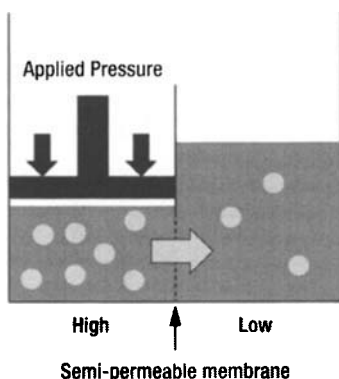


Figure 2.3 Reverse osmosis is the process by which an applied pressure, greater than the osmotic pressure, is exerted on the compartment that once contained the high-concentration solution, forcing water to move through the semipermeable membrane in the reverse direction of osmosis.

once contained the high-concentration solution (Figure 2.3). This pressure forces water to pass through the membrane in the direction reverse to that of osmosis. Water now moves from the compartment with the high-concentration solution to that with the low concentration solution. In this manner, relatively pure water passes through membrane into the one compartment while dissolved solids are retained in the other compartment. Hence, the water in the one compartment is purified or “demineralized,” and the solids in the other compartment are concentrated or dewatered.

Due to the added resistance of the membrane, the applied pressures required to achieve reverse osmosis are significantly higher than the osmotic pressure. For example, for 1,500 ppm TDS brackish water, RO operating pressures can range from about 150 psi to 400 psi. For seawater at 35,000 ppm TDS, RO operating pressures as high as 1,500 psi may be required.

2.3 Dead-End Filtration

The type of filtration illustrated in Figures 2.1, 2.2, and 2.3 is called “dead end” (“end flow” or “direct flow”) filtration. Dead end filtration involves all of the feed water passing through the membrane, leaving the solids behind on the membrane.

Consider a common coffee filter as shown in Figure 2.4. Feed water mixes with the coffee grounds on one side of the filter. The water then

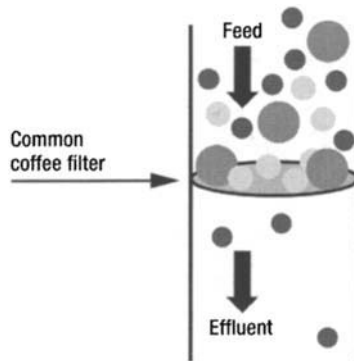


Figure 2.4 Dead-end filtration is a batch process that produces one effluent stream given one influent stream.

passes through the filter to become coffee that is largely free of coffee grounds. Virtually all of the feed water passes through the filter to become coffee. One influent stream, in this case water, produces, only one effluent stream, in this case coffee. This is dead end filtration.

Dead end filtration is a batch process. That means that the filter will accumulate and eventually blind off with particulates such that water can no longer pass through. The filtration system will need to be taken off line and the filter will need to be either cleaned or replaced.

2.4 Cross-Flow Filtration

In cross-flow filtration, feed water passes tangentially over the membrane surface rather than perpendicularly to it. Water and some dissolved solids pass through the membrane while the majority of dissolved solids and some water do not pass through the membrane. Hence, cross-flow filtration has one influent stream but yields two effluent streams. This is shown in Figure 2.5.

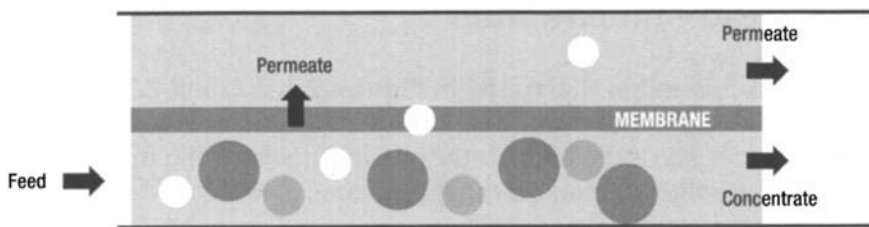


Figure 2.5 Cross-flow filtration is a continuous process that produces two effluent streams given one influent stream.

Cross-flow helps to minimize fouling or scaling of the RO membrane. In an effort to keep the membrane surface free of solids that may accumulate and foul or scale the membrane, tangential flow across the membrane surface aids in keeping the surface clean by scouring the surface; minimum flow rates across the membrane surface are required to effectively scour the surface. See Chapter 9.5 for more details about cross-flow filtration and RO system flow rates.

In theory, cross-flow is a continuous operation, as the scouring process keeps the membrane surface free of foulants. In practice, however, the scouring action of cross flow is not always enough to prevent all fouling and scaling. Periodically, the membranes will need to be taken off line and cleaned free of material that has accumulated at the surface.

Figure 2.6 is a simplified block diagram showing how cross-flow RO actually works. The diagonal line inside the rectangle represents the membrane. This diagram shows how the influent stream, with an applied pressure greater than the osmotic pressure of the solution, is separated into two effluent streams. The solution that passes through the membrane is called the permeate or product, and the solution retained by the membrane is called the concentrate, reject, waste, brine, or retentate. The flow control valve on the concentrate stream provides the back-pressure needed to cause reverse osmosis to occur. Closing the valve will result in an overall increase in pressure driving force, and a corresponding increase of influent water that passes through the membrane to become permeate.

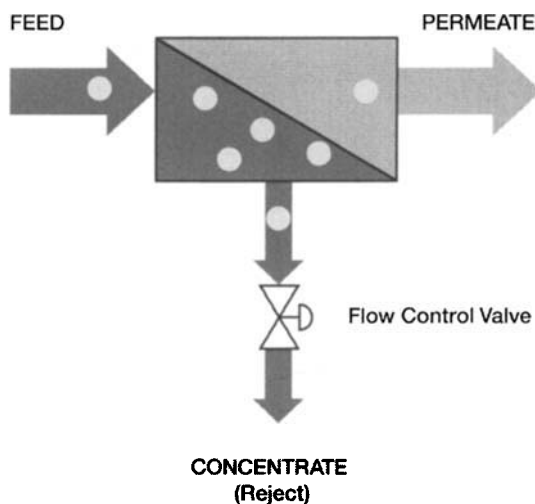


Figure 2.6 Cross-flow filtration showing concentrate flow control valve.

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Basic Terms and Definitions

This chapter defines basic terms used in conjunction with RO systems. Also see Chapter 9 for additional information as to how these parameters affect the performance of an RO system.

3.1 Reverse Osmosis System Flow Rating

An RO system is rated based on product flow rate. An 800-gpm RO would yield 800 gpm of permeate. The influent and reject flows are typically not indicated except in the design details (they are usually calculated knowing the product flow rate and the percent recovery).

In some cases, the actual design permeate flow rate of the RO system may differ from the “name plate” flow rating. In most of these situations, the RO system is de-rated by design due to a poor feed water source or as a natural result of low feed water temperature.

3.2 Recovery

Recovery (sometime referred to as “conversion”) is a term used to describe what volume percentage of influent water is “recovered” as permeate. Generally, RO *system* recoveries range from about 50% to 85%, with the majority of systems designed for 75% recovery. (Individual spiral wound membrane module recoveries vary from about 10% to 15%—see Chapter 4.3). A system recovery of 75% means that for every 100 gpm influent, 75 gpm will become permeate and 25 gpm will be retained as concentrate.

Recovery is calculated using the following equation:

$$\% \text{ Recovery} = (\text{permeate flow} / \text{feed flow}) * 100 \quad (3.1)$$

At 75% recovery, the concentrate volume is one-fourth that of the influent volume. If it were assumed that the membrane retains all the dissolved solids, they would be contained in one-fourth of the volume of influent water. Hence, the concentration of retained dissolved solids would be four times that of the influent stream

Table 3.1 Concentration factor as a function of recovery.

Recovery (%)	Concentration Factor
50	2
66	~ 3
75	4
80	5
83	6
87.5	8

(since not all dissolved solids are retained by the membrane, this becomes only an approximation). This is called the “concentration factor.” At 50% recovery, the concentrate volume would be one-half that of the influent water. In this case, the dissolved solids would be concentrated by a factor of two, so the concentration factor would be 2. Table 3.1 shows the concentration factor as a function of recovery. Understanding the reject concentration is important as the concentrate side of the membrane is the area where fouling and scaling occur (see Chapters 3.6 and 3.7).

Higher recovery results in the need to dispose of less reject water. However, higher recovery also results in lower-purity permeate. Consider the example shown in Figure 3.1. At the influent end of the membrane, the influent concentration is 100 ppm, while the recovery is 0%, and the membrane passes 2% total dissolved solids (TDS) (see Chapter 3.3). The permeate right at this spot would be about 2 ppm. As the influent water passes across more and more membrane area, more water is recovered. At 50% recovery, the concentration factor is 2, so the influent water now has a concentration of about 200 ppm. The permeate water at this point would now have a concentration of 4 ppm. At 75% recovery, the concentration factor is 4, so the influent water now has a concentration of about 400 ppm. The permeate water at this point would have a concentration of 8 ppm. Hence, higher recovery results in lower product purity.

The designer of the RO system selects the recovery for the system; it is not a property of the membrane. The designer must consider the trade off between higher recovery resulting in less concentrate water to dispose of but also lower permeate purity.

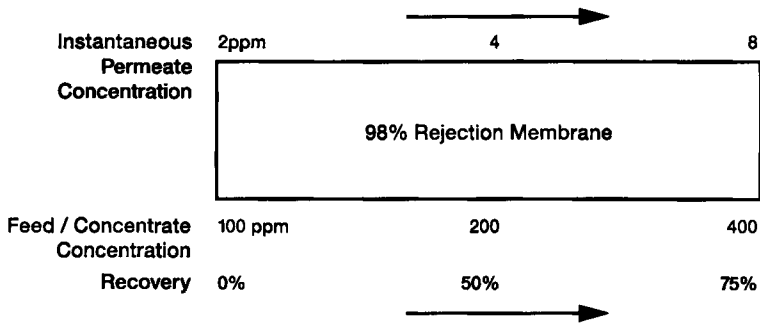


Figure 3.1 Concentrate and instantaneous permeate concentration as functions of recovery.

In practice, the recovery of the RO system is adjusted using the flow control valve located on the RO concentrate stream (see Figure 2.6). Throttling the valve will result in higher operating pressure, which forces more water through the membrane as opposed to down along the feed/concentrate side of the membrane, and results in higher recovery.

The recovery of an RO system is fixed by the designer. Exceeding the design recovery can result in accelerated fouling and scaling of the membranes, because less water is available to scour the membrane on the concentrate side. Falling below the design recovery will not adversely impact membrane fouling or scaling, but will result in higher volumes of wastewater from the RO system.

3.3 Rejection

Rejection is a term used to describe what percentage of an influent species a membrane retains. For example, 98% rejection of silica means that the membrane will retain 98% of the influent silica. It also means that 2% of influent silica will pass through the membrane into the permeate (known as "salt passage").

Rejection of a given species is calculated using the following equation:

$$\% \text{ Rejection} = [(C_f - C_p) / C_f] * 100 \quad (3.2)$$

where:

C_f = influent concentration of a specific component

C_p = permeate concentration of a specific component

Note that for exact calculation, the average feed concentration that takes in account both the feed and concentrate concentration rather than just the feed concentration at a single point in time should be used.

Salt passage is essentially the opposite of rejection:

$$\% \text{ Salt Passage} = 100 - \% \text{ Rejection} \quad (3.3)$$

$$\% \text{ Salt Passage} = (C_p / C_f) * 100 \quad (3.4)$$

Rejection is a property of the specific feed water component and the membrane of interest. Table 3.2 lists the general rejection ability of the most common polyamide composite RO membranes. Note that ionic charge of the component of interest plays a role its rejection by an RO membrane; the rejection of multi-valent ions is generally greater than for mono-valent ions.

In addition to the ionic charge, rejection of a particular species is also based on the following characteristics:¹

- Degree of dissociation: in general, the greater the dissociation, the greater the rejection, for example, weak acids are rejected better at higher pH.
- Molecular weight: in general, the greater the molecular weight, the greater the rejection, for example, the rejection of calcium is marginally better than the rejection of magnesium.
- Polarity: in general, the greater the polarity, the lower the rejection, for example, organics are rejected better than water.
- Degree of hydration: in general, the greater the degree of hydration, the greater the rejection, for example, chloride is rejected better than nitrate.
- Degree of molecular branching: in general, the more branching, the greater the rejection, for example, isopropanol is rejected better than normal propanol.

The rejection of gases is 0%, meaning that the concentration in the permeate stream will be the same as it is in the influent and concentrate streams. Gases that are not rejected include free chlorine that may be used to disinfect RO feed water through the pretreatment system (see Chapter 8.2) and carbon dioxide. RO sys-

Table 3.2 General rejection capabilities of most polyamide composite membranes at room temperature.

Species	Rejection (%)
Sodium	92 – 98
Calcium	93 – 99+
Magnesium	93 – 98
Potassium	92 – 96
Iron	96 – 98
Manganese	96 – 98
Aluminum	96 – 98
Ammonium*	80 – 90
Copper	96 – 99
Nickel	96 – 99
Zinc	96 – 98
Silver	93 – 96
Mercury	94 – 97
Hardness	93 – 99
Chloride	92 – 98
Bicarbonate	96 – 99
Sulfate	96 – 99+
Fluoride	92 – 95
Silicate	92 – 95
Phosphate	96 – 98
Bromide	90 – 95
Borate	30 – 50
Chromate	85 – 95
Cyanide	90 – 99+

* below pH 7.8. Above this pH, ammonia exists as a gas that is not rejected by RO membranes.

tems operating at near neutral pH will have some carbon dioxide in the feed water. Since gases are not rejected by RO membranes, the permeate and concentrate streams will also contain carbon dioxide. If the permeate is sent to ion exchange demineralization or electrodeionization after the RO, the carbon dioxide will use sites on the anion resin so that other anions are not well removed. In these cases, caustic soda (NaOH) is sometimes added to the RO feed water. This raises the pH and converts the carbon dioxide, which is not rejected by the RO membrane, to bicarbonate, which is rejected by the RO membrane. Caustic addition is recommended after sodium softening, which removes hardness (calcium, magnesium, barium, and strontium). Without softening, hardness in the feed water would saturate at the higher pH following caustic addition and scale the membranes. Caustic is also sometimes added between passes in a two-pass RO system (see Chapter 5.3; the first-pass RO removes the hardness while the effluent from the second pass is relatively free of carbon dioxide).

3.4 Flux

Flux is defined as the volumetric flow rate of a fluid through a given area. In the case of RO, the fluid is water and the area is that of the membrane. In the language of RO, flux is expressed as gallons of water per square foot of membrane area per day, (gfd). The flux of water through an RO membrane is proportional to the net pressure driving force applied to the water (see Chapter 4.1 for a discussion on transport models):

$$J = K (\Delta P - \Delta \Pi) \quad (3.5)$$

where:

J = water flux

K = water transport coefficient = permeability / thickness of the membrane active layer

ΔP = pressure difference across the membrane

$\Delta \Pi$ = osmotic pressure difference across the membrane

Note that the water transport coefficient is unique to a given membrane and is not a constant; it varies directly with temperature. The coefficient for some newer polyamide membranes also varies with pH.

The designer of the RO system chooses the flux rate; it is not a property of the membrane. In general, the flux that an RO system is designed for should be a function of the influent water quality. This is

Table 3.3 Recommended flux as a function of influent water source.

Feed Water Source	SDI	Recommended Flux, gfd *
RO Permeate	< 1	21 – 25
Well Water	< 3	14 – 16
Surface Supply	< 3	12 – 14
Surface Supply	< 5	10 – 12
Secondary Municipal Effluent— Microfiltration Pretreatment**	< 3	10 – 14
Secondary Municipal Effluent— Conventional Pretreatment	< 5	8 – 12

* for 8-inch diameter, brackish water membrane modules

** Microfiltration pore size < 0.5 microns.

because higher flux results in more rapid fouling of the membranes. So, the lower the influent water quality, the lower the operating flux of the RO system should be. Table 3.3 shows the recommended flux as a function of influent water source (which is an indirect measure of the water quality) and silt density index (SDI), which is a measure of the tendency of water to foul a membrane (See Chapter 3.9). When in doubt, a default flux of 14 gfd is usually recommended.

Specific flux is sometimes discussed in comparing the performance of one type of membrane with another. Specific flux is approximated by taking the overall system flux and dividing by the applied driving pressure:

$$\text{Specific Flux} = \text{Flux} / \text{Applied Pressure} \quad (3.6)$$

In comparing membranes, the higher the specific flux the lower the driving pressure required to operate the RO system. Specific flux is also defined as the permeability of the membrane.

3.5 Concentration Polarization

In simplest terms, the flow of water past an RO membrane is similar to that of the flow of water through a pipe, Figure 3.2. The flow in the bulk solution is convective, while the flow in the boundary layer is diffusive and is perpendicular to the convective flow of the bulk solution. There is no convective flow in the boundary layer.

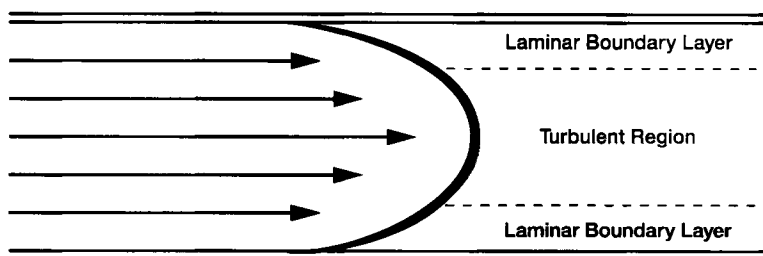


Figure 3.2 Hydraulic boundary layer formed with fluid flow in a pipe.

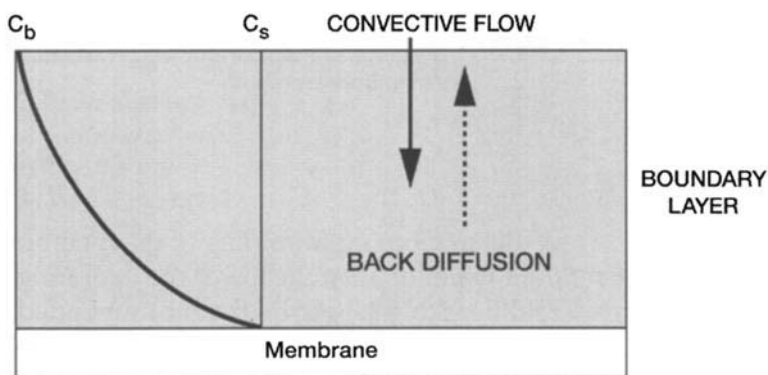


Figure 3.3 Concentration polarization, where C_b is the bulk concentration and C_s is the concentration at the membrane surface.

The slower the velocity of water through the pipe, the thicker the boundary layer becomes.

Now, consider flow along the surface of a membrane. The same boundary layer forms as with flow through a pipe. However, with a membrane system, because there is a net flow out through the membrane, there is convective flow to the membrane, but only diffusional flow away from the membrane. Since diffusion is slower than convection, solutes rejected by the membrane tend to build up on the surface and in the boundary layer. Thus, the concentration of solutes at the membrane surface is higher than in the bulk solution. This boundary layer is called “concentration polarization.”² The phenomenon is shown in Figure 3.3.

Concentration polarization has a negative effect on the performance of an RO membrane. It acts to reduce the throughput of the membrane in three important ways. First, it acts as a hydraulic resistance to water flow through the membrane. Second, the build up of solutes increases

the osmotic pressure within the boundary layer, effectively reducing the driving force for water through the membrane. Third, the higher concentration of solutes on the membrane surface than in the bulk solution, leads to higher passage of solutes than would be predicted by the feed water concentration. This is because an RO membrane rejects solutes based on the concentration of the salt that it “sees.” If the concentration of a species is higher at the membrane surface, as is the case with concentration polarization, the amount of solute passing into the permeate will be higher than the expected amount of solute based on the bulk concentration of that solute. The actual rejection and salt passage exhibited by the membrane does not change. However, the apparent rejection/passage does. For example, assume that the bulk concentration of silica is 10 ppm, while the concentration at the membrane surface is 11.5 ppm. If the rejection is 98%, the silica concentration that would be expected to be in the permeate based on the bulk concentration is 0.20 ppm. However, the membrane “sees” 11.5 ppm, so the actual salt passage is 2% of 11.5 ppm, or 0.23 ppm. Actual rejection is still 98%. Apparent rejection is 97.7%.

(See reference 1 for a more complete discussion about concentration polarization.)

3.6 Beta

Beta, sometimes called the “concentration polarization factor,” is the ratio of the concentration of a species at the membrane surface to that in the bulk solution. Hence, Beta is a way of quantifying concentration polarization.

The higher the Beta number, the more likely the membranes are to foul or scale. Again, since Beta measures the ratio of concentration at the surface to that in the bulk solution, the higher the beta number, the higher the relative concentration at the surface. If the concentration at the surface gets high enough, saturation may be reached and scale will begin to form. Maximum acceptable Beta typically ranges from about 1.0 to 1.2 to minimize formation of scale. (see Chapter 9.6).

Beta is not a property of the membrane; it is an artifact of the system design that is selected. Specifically, Beta is a function of how quickly the influent stream is dewatered through the RO system. If water is removed too quickly from the influent stream, Beta will increase, as a relatively high volume of dissolved solids is left behind on the membrane because of the high volume of water that permeates

through the membrane. Concentration polarization further exacerbates the problem because of the diffusional-only flow away from the membrane surface. See Chapter 9.6 for more information about Beta and its relationship with water flux and salt passage.

3.7 Fouling

Membrane fouling is a result of deposition of suspended solids, organics, or microbes on the surface of the membrane, typically on the feed/concentrate side. Fouling species include:

- colloids, such as alumina- and iron-silicates. Silica can precipitate at concentration below saturation in the presence of aluminum or iron.¹
- organics, which provide nutrients for microbes,
- microbes,
- color, which irreversibly adsorbs onto the membrane polymer,
- metals, such as iron and manganese that precipitate when oxidized; aluminum typically from alum, which is commonly overfed, particularly into municipal/surface sources; and hydrogen sulfide, which releases elemental sulfur upon oxidation, a sticky material very difficult if not impossible to remove from a membrane.

Table 3.4 lists generally-accepted water quality guidelines to minimize fouling of RO membranes.³

Table 3.4 Generally-accepted water quality guidelines to minimize RO membrane fouling.

Species	Measure	Value
Suspended Solids	Turbidity	< 1 NTU
Colloids	SDI	< 5
Microbes	Dip Slides*	< 1,000 CFU/ml**
Organics	TOC Concentration	< 3 ppm
Color	Color units	< 3 APHA
Metals: iron, manganese, aluminum	Concentration	< 0.05 ppm
Hydrogen Sulfide	Concentration	< 0.1 ppm

* Other methods see Chapter 7.3

** In RO reject stream

Fouling is exacerbated by high membrane flux and low cross-flow velocity, both conditions that increase concentration polarization. Higher flux translates into water being removed through the membrane at a faster rate, leaving behind solids that now accumulate more rapidly in the concentration polarization boundary layer. If the residence time is sufficient in the boundary layer, these solids will deposit on the membrane surface, sometimes permanently. Cross-flow velocity affects the thickness of the boundary layer. Lower cross-flow velocity corresponds to a thicker boundary layer. A thicker boundary layer allows for greater accumulation of solids in the layer, and the solids spend more time in the layer due to the increased thickness of the boundary layer, setting up the potential for accelerated fouling of the membrane.

A fouled membrane exhibits two major performance issues: higher than normal operating pressure (to compensate for lower membrane flux at constant pressure) and higher than normal pressure drop. As foulants collect on the surface of the membrane, they form an additional barrier layer to transport through the membrane (see Figure 3.4). This additional barrier requires a greater net driving force, which translates into higher operating pressure. Higher pressure drop is a result of the increased resistance to cross-flow caused by the layer of foulants. Pressure drop translates into axial pressure on the membrane module (see Chapter 11.3.1.3). If the pressure drop gets high enough, the axial pressure on the membrane module can

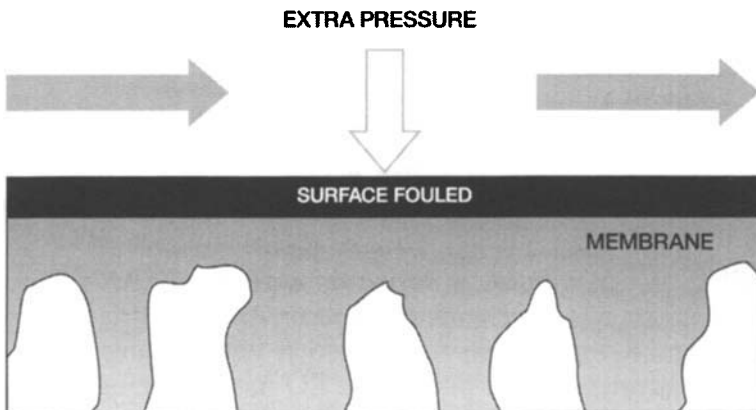
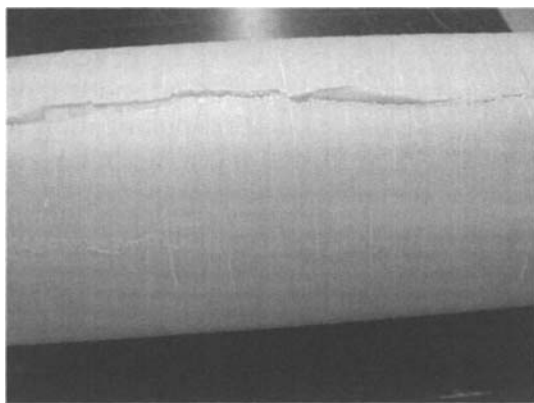


Figure 3.4 Fouling on membrane surface creates an additional barrier to permeate transport that requires additional pressure to force permeate through the fouling layer.

(a)



(b)

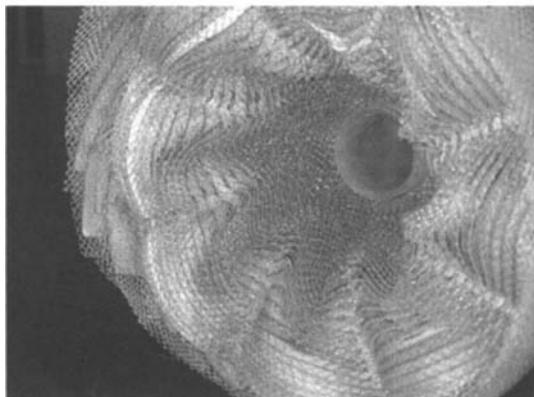


Figure 3.5 Cracked outer module casing (a) and telescoped membranes and spacers (b) due to excessive pressure drop.

become great enough to cause the membrane and module to fail (see Figures 11.5 and 11.6). Failure can manifest as cracks in the outer module casing and telescoped membranes and spacers (Figure 3.5A and B respectively).

Fouling of a membrane is primarily a physical filtration event, although in some cases, charge of a species can determine its potential to foul an RO membrane (as is the case with cationic coagulants on the negatively-charged polyamide RO membrane—see Chapter 8.1.1.1). The lead membrane modules in an RO system are generally most affected by fouling (“lead” membrane module is the end module in a pressure vessel that feed water contacts first). One exception is

microbial fouling. Microbes can establish colonies anywhere in the membrane system where conditions favor growth. Satellite colonies can break off and further distribute themselves throughout the system. Note that even short term exposure to potential foulants can result in long-term and even permanent fouling issues for a membrane.⁴

3.8 **Scaling**

Scaling of RO membranes is a result of precipitation of saturated salts onto the surface of the membrane. Table 3.5 lists generally-accepted water guidelines for minimizing scaling of RO membranes.³ The table includes the following species:

- calcium scales, including carbonate, sulfate, fluoride, and phosphate,
- reactive silica, which is measured in the RO reject and is a function of temperature and pH, and
- sulfate-based scales of trace metals, such as barium and strontium.

Scaling is exacerbated by high membrane flux and low cross-flow velocity, in the same manner as membrane fouling is increased. Higher flux brings more solutes into the concentration polarization boundary layer quicker. If the concentration of the solutes in the boundary layer reaches saturation, these solutes will scale the membrane. Lower cross-flow velocity corresponds to a thicker boundary layer. This increases the residence time for solute within the boundary layer, increasing the chance that saturation will be achieved and scale will form.

Table 3.5 Generally-accepted water quality guidelines to minimize RO membrane scaling.

Species	Measure	Value
Silica (Soluble)	ppm	200*
Barium, Strontium	ppm	< 0.05
Calcium	LSI	< 0**

* In RO reject stream

** Can go up to 2.0 – 2.5 with appropriate antiscalant

A scaled membrane exhibits three major performance issues: higher than normal operating pressure (to compensate for lower flux at constant pressure), higher pressure drop, and lower than expected salt rejection. As with fouled membranes, scale forms an additional barrier layer through which water has to travel. Additional driving force is required to push water through this barrier layer if the same productivity is to be maintained. Higher pressure drop is due to resistance of the scale to cross flow. Lower salt rejection is a function of concentration polarization, in that the concentration of the scaled mineral is higher at the membrane surface than in the bulk solution. Thus, the membrane “sees” a higher concentration, and, although the intrinsic rejection by the membrane remains constant, the actual amount of a solute that passes through the membrane is greater (see Equation 4.2, which predicts that the flux of a solute through an RO membrane is a function of the concentration difference between the solute in the boundary layer and in the permeate). Hence, the apparent rejection is lower and product purity is lower.

Since scaling is a concentration phenomenon, it goes to reason that scale would be most likely found in the last stages of an RO system where the concentration of salts is the highest. To determine the potential for a salt to form scale, the ion product of the salt in question (taken in the reject stream) is compared with the solubility product for the salt under the conditions in the reject.

Equation 3.7 defines the ion product at any *degree of saturation*:

$$IP = [\text{cation}]^a[\text{anion}]^b \quad (3.7)$$

Where:

IP = ion product

[cation] = cation concentration

[anion] = anion concentration

Superscripts:

a = the quantity of cation present within the salt

b = the quantity of anion present within the salt.

Equation 3.8 defines the solubility product at *saturation*:

$$K_{sp} = [\text{cation}]^a[\text{anion}]^b \quad (3.8)$$

where:

K_{sp} = solubility product

[cation] = cation concentration

[anion] = anion concentration

Superscripts:

a = the quantity of cation present within the salt

b = the quantity of anion present within the salt.

In general, scale will form when the ion product is greater than the solubility product. For sulfate-based scales, scaling can occur when the ion product is greater than 80% of the solubility product.¹

Scaling indexes are used to aid in the determination of whether a salt will scale an RO membrane. The Langelier Saturation Index (LSI) and the Stiff-Davis Saturation Index are commonly used to predict calcium carbonate scale (see Chapter 3.10). Design programs developed by membrane manufacturers typically report scaling indexes for other forms of scale as a percent of saturation (see Chapter 10). A scaling index is 100% when the ion product equals the solubility constant for most salts. For sulfate-based scales, the saturation index is 100% when the ion product is equal to 80% of the solubility constant. Feed water softening or the use of antiscalants is required when the saturation index is 100% or greater.

3.9 Silt Density Index

Silt density index (SDI) is a test that measures the potential of an influent water to foul an RO membrane with suspended solids and colloids. The test involves timing flow of the influent water through a 0.45-micron filter pad.⁵

Figure 3.6 shows the basic materials required to run an SDI test. The apparatus shown in Figure 3.6 should be connected as close to the inlet of the RO as possible (preferably between the cartridge filters and the RO, if possible).

Procedures to run an SDI test call for a 0.45-micron filter pad to be placed with the shiny side up in the filter holder. A small squirt of water from a water bottle is used to wet and completely seat the filter pad. (If the pad shows signs of air bubbles or erratic coverage of foulant of the pad upon removal from the filter holder, the filter was not seated properly and the SDI test is invalid). The isolation valve is opened and the pressure regulator is set to 30 psi. The time it takes to collect 500 ml of water through the 0.45-micron filter pad is then recorded. After this volume of water has been collected, water is allowed to run through the filter continuously for a total time of 15 minutes. If necessary, the pressure should be adjusted to remain at 30 psi. At the end of 15 minutes, the time it takes to collect another

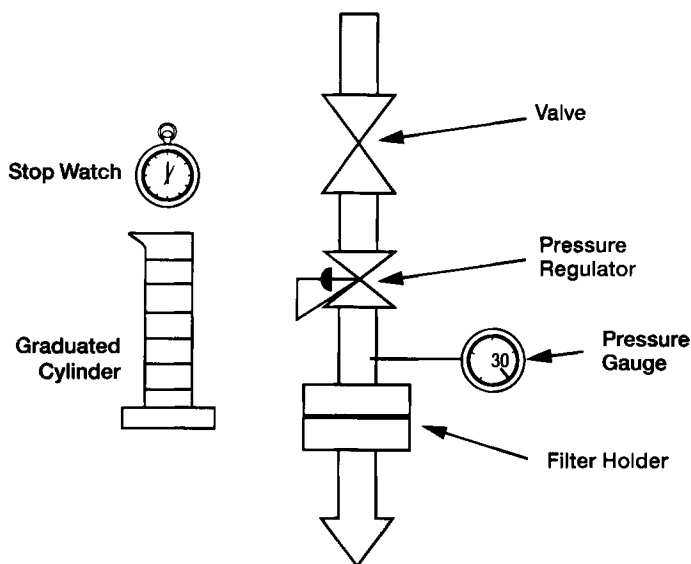


Figure 3.6 Silt density index test apparatus and ancillary equipment.

500 ml of water is recorded. Both recorded times are then entered into Equation 3.9 to calculate the SDI (note that SDI is unit-less):

$$SDI_n = \frac{1 - t_0/t_n}{n} \times 100 \quad (3.9)$$

where:

n = total run time, 15 minutes

t_0 = time to collect 500 ml of influent water at time = 0, in seconds

t_n = time to collect 500 ml of influent water at time = n (15 min), in seconds

The test must be run at 30 psi. As mentioned above, if the pressure drops during the test, it should be adjusted back to 30 psi.

(Note that the SDI test can also be run for 5 minutes and 10 minutes. However, most literature references to SDI discuss the 15 minute-run, unless otherwise stated.)

The maximum SDI_{15} that can possibly be calculated from equation 3.9 occurs when the time to collect the final 500-ml of water, t_n , is infinite. The maximum SDI_{15} then becomes $(1/15) \times 100$ or 6.7.

An SDI test must be conducted on line using a representative sample of feed water; shipping samples off to a laboratory for testing

is not recommended, as solids may settle or adhere to the shipping container, invalidating the test. Silt Density Index is preferably run as close as possible to the influent of the RO membranes, which means after the cartridge filter. If testing after the cartridge filter is not possible, then just before the cartridge filter is the next best location. Water should be run through the sample port or tap for at least an hour and preferably overnight before running an SDI test. This is necessary to realize a representative sample of feed water that is free of solids that may have accumulated over time in the sample port or tap.

Although there are no truly automated SDI devices, there are semi-automatic devices in addition to the manual device shown in Figure 3.6. These semi-automatic units run the SDI test, including the timing and collection of water through the SDI filter. Human intervention is required to replace the SDI filter pad and to record the test results. Note that the automatic units should be flushed with high-quality water after each use, as there are “dead” spots in the lines where contaminants and bacteria can collect and foul the tubing. Without flushing, these contaminants will affect subsequent SDI tests.

The lower the SDI, the lower the potential for fouling a membrane with suspended solids. Membrane manufacturers require that the SDI_{15} be less than 5 to meet their warranty conditions. However, even though the SDI may be less than 5, the water may still be capable of fouling the membranes. In practice, an SDI_{15} of less than 3 is usually necessary to reduce the potential for fouling to an acceptable level.

Note that SDI and turbidity are only loosely related. In general, the higher the turbidity, the higher the SDI will be. However, low turbidity (<1 NTU) does not, in turn, imply low (<5) SDI.

Figure 3.7 shows SDI filter pads taken before and after a multimedia filter treating RO influent water. The pads in this figure illustrate 2 important issues. First, the filter pads provide a visual confirmation about the efficacy of the multimedia filter to reduce the concentration of suspended solids in the RO influent water. Second, the filters pads can be analyzed to determine the nature of the deposit on them. Visually, the following colors are indicative of specific potential foulants:

- Yellow: possibly iron or organics
- Red to reddish brown: iron
- Black: manganese (if the color dissolves when the pad is treated with acid)

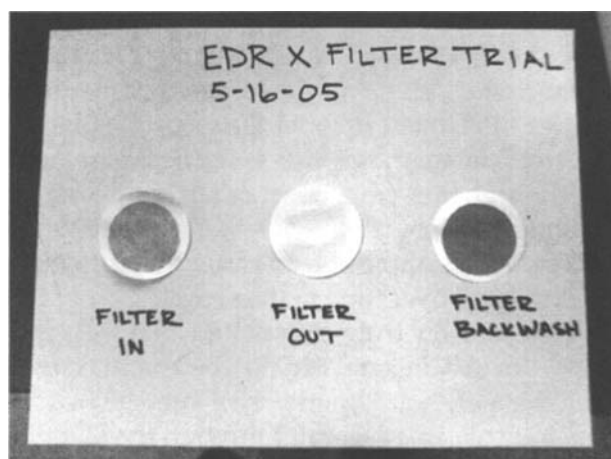


Figure 3.7 Silt density index pads taken before and after a filter treating RO influent water.

The SDI filter pads can also be sent into a lab for analysis of the deposit. The results of the deposit analysis will aid in the development of an appropriate pretreatment scheme, as specific species contributing to the suspended solids loading can be targeted for treatment to reduce their concentration in the feed water to the RO.

3.10 Langelier Saturation Index

The Langelier Saturation Index (LSI) is a method for quantifying the scaling or corrosion tendency of water. It was originally applied to cooling water. The LSI is based on the pH and temperature of the water in question as well as the concentrations of TDS, calcium hardness, and alkalinity.

Langelier Saturation Index is calculated using equation 3.10:

$$\text{LSI} = \text{pH} - \text{pH}_a \quad (3.10)$$

where:

$$\text{pH}_a = (9.30 + A + B) - (C + D) \quad (3.11)$$

and:

$A = (\text{Log}_{10}[\text{TDS}] - 1)/10$, where [TDS] is in ppm

$B = -13.12 \times \text{Log}_{10}(\text{°C} + 273) + 34.55$

$C = \text{Log}_{10}[\text{Ca}^{2+}] - 0.4$, where $[\text{Ca}^{2+}]$ is in ppm CaCO_3

$D = \text{Log}_{10}[\text{alkalinity}]$, where [alkalinity] is in ppm CaCO_3

Table 3.6 Langelier saturation index.

LSI	Condition
3.0	Extremely severe scaling
2.0	Very severe scaling
1.0	Severe scaling
0.5	Moderate scaling
0.2	Slight scaling
0.0	Stable water (no scale)
-0.2	No scale, very slight tendency to dissolve scale

If LSI is greater than 0, the water has the tendency to form calcium carbonate scale. If the LSI is equal to 0, the water is in chemical balance. If the LSI is less than 0, the water is corrosive (refer to Table 3.6).

Langelier Saturation Index is valid up to about 4,000 ppm TDS. At higher TDS concentrations, the Stiff-Davis Saturation Index (SDSI) is used, Equation 3.12:

$$\text{SDSI} = \text{pH} - \text{pCa} - \text{p}_{\text{ALK}} - K \quad (3.12)$$

where:

$\text{pCa} = -\text{Log}_{10}[\text{Ca}^{2+}]$, where $[\text{Ca}^{2+}]$ is in ppm

$\text{p}_{\text{ALK}} = -\text{Log}_{10}[\text{total alkalinity}]$, where alkalinity is in ppm

K = a constant based on the total ionic strength and temperature

For RO applications, a positive LSI or SDSI indicates that the influent water has a tendency to form calcium carbonate scale. In these cases, pretreatment in the form of softening (either with lime or ion exchange), or via the use of antiscalants and/or acid is required.

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Membranes

Membranes

Pertinent to the understanding of the operation of an RO system is the fundamental knowledge of various theoretical models describing movement of solutes and water through an RO membrane. By understanding how solutes and water are transported through membranes, appropriate modifications can be made to the membrane polymers to improve performance (flux and rejection). See the book by Richard Baker, *Membrane Technology and Applications*, 2nd edition (John Wiley & Sons, 2004) for more detail about the history and development of membrane and transport models.

This chapter covers the development of transport models. Also discussed are basic membrane polymers and modules, and how each is made.

4.1 Transport Models

The purpose of a transport model is to mathematically relate performance (typically flux (see Chapter 3.4) of both solvent and solute) to operating conditions (typically pressure and concentration driving forces).¹ The objective is to predict membrane behavior under certain conditions.

There are several models that describe the transport of mass through RO membranes. These are based on different assumptions and have varying degrees of complexity. The solution-diffusion model best describes the performance of “perfect,” defect-free membranes and is considered the leading theory on membrane transport.² Three other theories are presented here for completeness.

Transport models fall into three basic classifications: models based on solution/diffusion of solvents (nonporous transport models), models based on irreversible thermodynamics, and models based on porous membranes. Highlights of some of these models are discussed below.

4.1.1 Solution-Diffusion Model (non-porous model)

The solution-diffusion transport model was originally described by Lonsdale et. al.³ This model assumes that the membrane is nonporous (without imperfections). The theory is that transport through the membrane occurs as the molecule of interest *dissolves* in the membrane and then diffuses through the membrane. This holds true for both the solvent and solute in solution.

In the solution-diffusion model, the transport of solvent and solute are independent of each other, as seen in Equations 4.1 and 4.2. The flux of solvent through the membrane is linearly proportional to the effective pressure difference across the membrane (Equation 4.1):

$$J_w = A(\Delta P - \Delta \Pi) \quad 4.1$$

where:

J_w = flux of solvent

A = water permeability coefficient (a function of the diffusivity of water through the membrane)

ΔP = applied pressure driving force (a function of the feed, concentrate, and permeate pressures)

$\Delta \Pi$ = osmotic pressure of the solution (a function of the feed, concentrate, and permeate concentrations)

The flux of solute through the membrane is proportional to the effective solute concentration difference across the membrane (Equation 4.2):

$$J_s = K(C_{A2} - C_{A3}) \quad 4.2$$

where:

J_s = flux of solute

K = salt permeability coefficient (a function of the salt diffusivity through the membrane)

C_A = molar concentration of solute

Subscripts:

2 = at the boundary layer

3 = in the permeate

Equations 4.1 and 4.2 are most commonly used to describe transport of water and solutes through membranes due to their simplicity

and their close approximation to empirical data. Consider Figure 4.1, which shows the flux of both water and salt and the rejection of salt exhibited by a seawater membrane as a function of applied pressure². Specifically, a seawater FilmTec FT-30 membrane was operated on a 35,000 ppm sodium chloride solution with an osmotic pressure of 350 psi (2.5 MPa). As the figure shows, there is virtually no water flux until the applied pressure exceeds the osmotic pressure of the solution. Upon commencement of water flux, it increased linearly with increasing pressure, as is predicted by Equation 4.1. The salt flux, on the other hand, remained fairly constant over the range of applied pressure, as predicted by equation 4.2. Hence, as applied pressure is increased, progressively more water passes through the membrane relative to salt. This leads to the conclusion

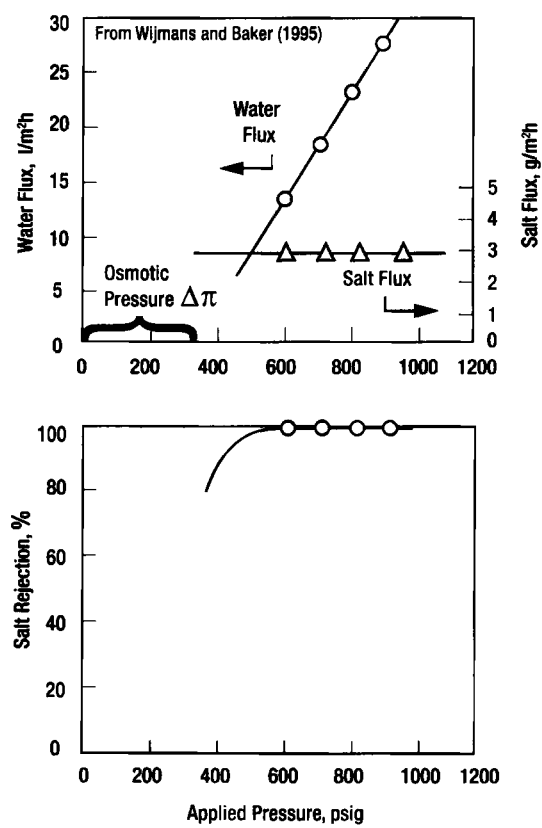


Figure 4.1 Flux and rejection data for a seawater FilmTec FT-30 membranes operating on 35,000 ppm (350 psi osmotic pressure) sodium chloride solution.²

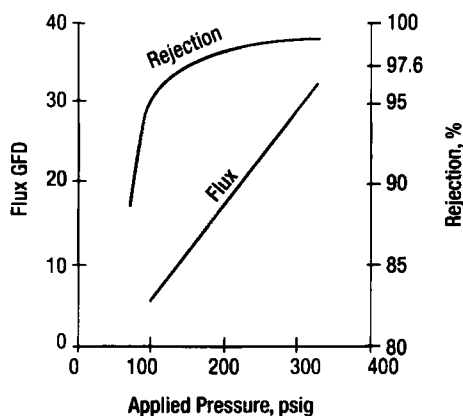


Figure 4.2 Polyamide membrane flux and rejection as a function of applied pressure. Test conditions: 5,000 ppm NaCl solution at 25°C.⁴

that the concentration of salt in the permeate should decrease with increasing applied pressure. Referring back to Equation 3.2, as the concentration in the permeate goes down, the percent salt rejection should approach 100% as applied pressure increases. Indeed, that is what is shown in Figure 4.1. These results are confirmed in Figure 4.2 for a low-pressure membrane operated on a 5,000 ppm sodium chloride solution at 25°C.

4.1.2 Solution – Diffusion Imperfection Model (porous model)

The solution–diffusion theory models the performance of the perfect membrane. In reality, industrial membranes are plagued with imperfections that some argue must be considered when developing a complete theory that models performance. The basis of the Diffusion Imperfection Model is the assumption that slight imperfections in the membrane occur during manufacturing that allow for leakage of solution through the membrane.⁵ This model helps explain why lower-than-projected separation of solutes and water were observed for industrial membranes than was predicted by the non-porous, solution – diffusion model.

Water flux through the membrane is represented by Equation 4.3. This flux is based on the solutions – diffusion model with the added term to reflect transport due to the imperfections.

$$N_w = J_w + K_3 \Delta PC_w = A(\Delta P - \Delta \Pi) + K_3 \Delta PC_w \quad 4.3$$

where:

N_w = total water flux

K_3 = coupling coefficient

C_w = concentration of water on the feed side of the membrane

The solute flux is given in Equation 4.4:

$$N_s = J_s + K_3 \Delta PC_R = K(C_{A2} - C_{A3}) + K_3 \Delta PC_R \quad 4.4$$

where:

N_s = total solute flux

C_R = solute concentration on the feed side of the membrane.

Again, the solute flux is equivalent to that for the solution – diffusion model (Equations 4.1 and 4.2) with the added term to represent the flow through the imperfections.

Experiments have shown that the solution – diffusion imperfection model fits data better than the solution – diffusion model alone and better than all other porous flow models.⁶ However, the solution-diffusion model is most often cited due to its simplicity and the fact that it accurately models the performance of the perfect RO membrane.

4.1.3 Finely-Porous Model (porous model)

The finely-porous model is based on a balance of applied and frictional forces in a 1-dimensional pore.⁷ The model considers friction between the solute and solvent, and between the solute and the membrane material. The model also includes the membrane thickness and the fractional pore area of the membrane surface.

Due to the complexity of the model, it is not represented mathematically here, but the reader is advised to consider references¹ and⁸ for further details.

4.1.4 Preferential Sorption – Capillary Flow Model (porous model)

This model is based on a generalized capillary flow model that includes viscous flow for water and solute transport, and for pore

diffusion.⁹ It further relies on film theory for transport through the boundary layers. The model states that by applying pressure, both the solvent *and* solute permeate through the micropores of the membrane, with water preferentially adsorbed onto the pore walls. Salt is rejected at the membrane surface for physiochemical reasons. Transport through the membrane is only through pores.

Solvent flux is given by Equation 4.1, where transport is proportional to the pressure driving force. The total solute flux depends on diffusion and is given by Equation 4.5:

$$N_s = (D_{AM} K / T) (C_{A2} - C_{A3}) \quad 4.5$$

where:

D_{AM} = diffusivity of solute in membrane

T = effective thickness of the membrane

C_{A2} = molar concentration of solute in boundary layer

C_{A3} = molar concentration of solute in permeate

4.1.5 Phenomenological Transport Relationship (Irreversible thermodynamics)

Phenomenological transport relationships can be developed even in the absence of any knowledge of the mechanisms of transport through the membrane or any information about the membrane structure.¹⁰ The basis of irreversible thermodynamics assumes that if the system is divided into small enough subsystems in which local equilibrium exists, thermodynamic equations can be written for the subsystems.

As with the finely-porous model, (Chapter 4.1.3), the mathematical representation of solvent and solute fluxes for the irreversible thermodynamic model is quite complex and beyond the scope of this work. However, it is recommended that readers consider references¹ and⁸ for details on this transport model.

4.2 Membrane Materials

The performance of reverse osmosis is directly dependent on the properties of the membrane material.¹¹ More specifically, the chemical nature of the membrane polymer and the structure of the membrane are what determines the rejection and flux properties

of the RO system. Ideally, RO membranes should offer high flux and high rejection, in addition to high strength and durability. In practice, however, high rejection and high flux have been two mutually-exclusive goals that have eluded researches for decades of membrane development. Although the last few years has seen an increase in flux rates with no decrease in rejection (and in some cases, a slight increase in rejection), most membranes today represent a compromise between high rejection and high flux.(1, 11)

Two most common families of RO membranes, based on the type of polymer backbone, are cellulose acetate and polyamide.¹² Membranes made from these polymers differ in many respects, including performance, physical properties, structure and the manner in which they are created. These aspects are discussed below.

4.2.1 Cellulose Acetate Membranes—Asymmetric membranes

Cellulose acetate (CA) membranes were the first commercially-viable RO membranes developed.^{13,14} These membranes were commercially viable because of their relatively high flux due to the extreme thinness of the membrane.¹⁵ High flux was important to reduce the size and cost of an RO system.

Early cellulose acetate membranes were created by Loeb and Sourirajan using the non-solvent phase separation or “phase inversion” method.¹⁶ In short, this method involves dissolving a non-water soluble polymer (cellulose acetate) in an organic solvent (such as acetone) along with a casting-solution modifier (initially magnesium perchlorate but later a swelling agent, formamide) and then casting a film on a sturdy surface such as a sheet of glass using a thin blade.(2, 17) The film is left to stand for 10–100 seconds to partially evaporate the solvent. As the solvent evaporates, there is an increase in concentration of polymer at the solution/air interface, since the solvent evaporates more rapidly from the surface. This results in two phases within the film: a polymer-rich phase and a polymer-poor phase. Before the solvent completely evaporates, the membrane is immersed in water to completely form a thin “skin” as the remaining polymer diffuses out. The membrane is then “annealed” (heated to 70–90°C) in a hot water bath, which forms the small voids in the polymer-poor phase behind the skin. This porous region becomes the support structure. The thickness of the skin is typically about 0.2 microns, while the thickness of the entire membrane is

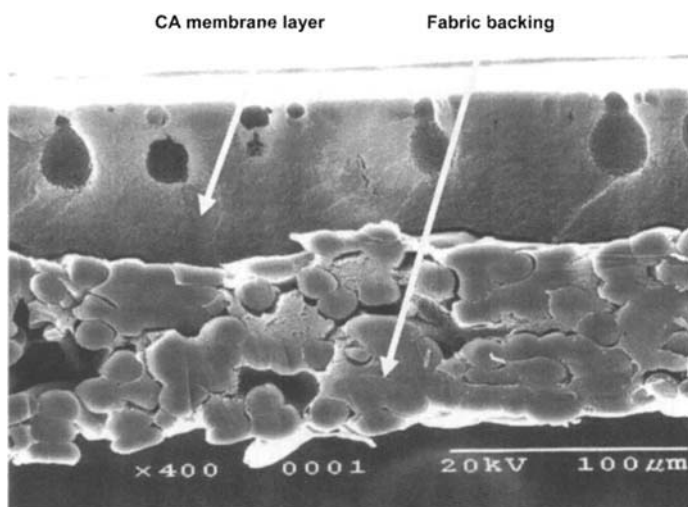


Figure 4.3 Cross section of a cellulose acetate RO membrane.

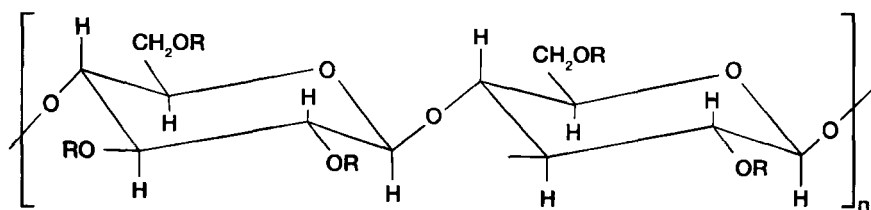


Figure 4.4 Chemical structure of cellulose acetate RO membranes.

about 100 microns.¹⁹ It is the extreme thinness of the skin and the relatively high void volume of the supporting structure that gave these early cellulose acetate membranes the high rejection and high flux needed for commercial viability. Subsequent development allowed for preparing the membrane from a blend of cellulose diacetate and triacetate and casting the membrane on a fabric backing.

Figure 4.3 shows a cross section of a CA membrane. The structure is asymmetric or “anisotropic,” with a solute-rejecting layer on top of a microporous support, all made of the one polymeric material.

Figure 4.4 shows the chemical structure of CA membranes. Because the functional groups at the ends of the polymer chains are not highly charged, the membrane itself is considered uncharged.¹² This is an important characteristic, especially if charged (cationic) polymers are used ahead of the membrane to pretreat the influent water. The

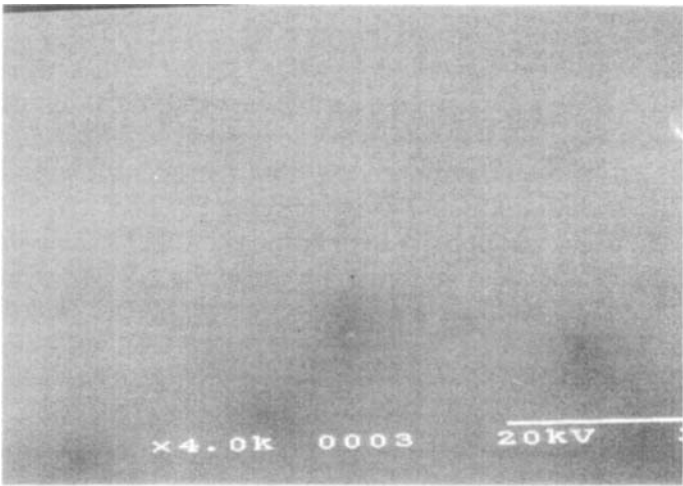
neutrality of charge will serve to minimize the potential for fouling with any polymer that carries over from the pretreatment system.

Table 4.1 lists the predominant characteristics of a CA membrane. Some of these characteristics are advantageous while others present quite severe limitations in using this type of membrane.

- The relatively smooth surface morphology of the membrane offers some protection from fouling, in that there are no obvious dead spaces on the membrane where foulants can become trapped. See Figure 4.5.
- Neutral surface charge minimizes the potential for fouling with cationic polymer that might carry over from the pretreatment equipment.
- Being able to tolerate up to 1 ppm free chlorine on continuous basis offers some protection from biological growth on the membrane. This is particularly important because the CA polymer itself supplies nutrients for microbial populations, which then metabolize the polymer and degrade the membrane.
- Temperature is limited to 35°C. Higher temperatures will further anneal the membrane, resulting in a much denser material that is difficult to force water through. Hydrolysis of the membrane may also occur, also leading to degradation of the membrane.¹⁹
- Operating pH of a CA membrane is limited to 4–6. This implies that acid is required to drop the influent pH to about 5 to 5.5 during nominal operations such that the reject pH remains less than 6. pH extremes will cause hydrolysis of the membrane polymer. Figure 4.6 shows lifetime of a CA membrane as a function of pH.
- Operating pressure for a CA membrane ranges from about 200 – 400 psig. The comparatively high pressure for brackish water applications is necessary because of the relatively dense nature of the CA membrane. Note that pressures above 400 psig (2.8MPa) will result in compaction of the membrane. Compaction occurs when the pressure essentially “crushes” the membrane, making it even denser.
- Salt rejection varies depending on whether “standard” or “high rejection” CA membranes are used. Rejection of silica, however, is only about 80%.

Table 4.1 Characteristics of cellulose acetate RO membranes.

Property	Value for CA Membranes
Membrane Type	Homogenous asymmetric
Salt Rejection (%)	~95
Silica Rejection (%)	~85
pH Range	4–6
Feed Pressure (brackish membrane)	220–440 psi
Temperature Tolerance	Up to 30°C
Surface Charge	Neutral
Chlorine Tolerance	Up to 1 ppm continuously
Biological growth	Metabolized membrane
Fouling Tolerance	Good
Surface Roughness	Smooth



SEM picture of surface of cellulose acetate membrane

Figure 4.5 Smooth surface morphology of cellulose acetate membranes.

- Hydranautics makes a commonly used “CAB” cellulose acetate membrane.

Cellulose acetate membranes were originally cast in tubular form (refer to Figure 1.3). These tubular membranes were used in

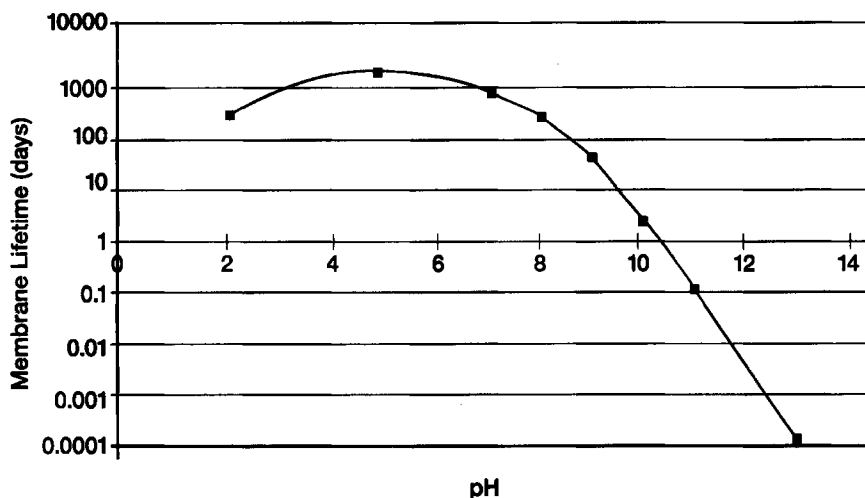


Figure 4.6 Lifetime of a CA membrane as a function of pH.

the first commercial RO system at Coalinga, California in 1965. The membranes were developed and prepared at UCLA (see Chapter 1). The break for commercial viability came when as an appropriate swelling agent, formamide, was found for the cellulose acetate membrane during preparation.¹⁷

4.2.2 Polyamide and Composite Membranes

Polyamide membranes were developed in an effort to improve upon the performance of CA membranes. In particular, the higher operating pressure and relatively low salt rejection of CA membranes were holding back RO technology from becoming more commercially viable.

4.2.2.1 Linear Aromatic Polyamide Membranes

Aromatic polyamide membranes were developed by a few companies, including Toray, Monsanto and DuPont. DuPont developed a linear aromatic polyamide (nylon) membrane with pendant sulfonic acid groups, which they commercialized as the PermasepTM B-9 and B-10 membranes and is shown in Figure 4.7 (Permasep is a registered trademark of E. I. Du Pont De Nemours & Company, Inc. Wilmington, DE). Just as CA membranes were created out of

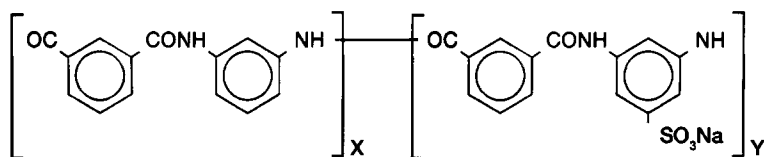


Figure 4.7 Aromatic polyamide membrane developed by DuPont.

a single polymer, Permasep membranes were also created out of single polymer rather than a composite (see below). These membranes exhibited higher rejection capabilities at higher flux and somewhat lower operating pressures than CA membranes. They were originally fabricated into hollow fine fiber membranes and used primarily for seawater and brackish water desalination and some specialty applications such as recovery of electric deposition paints (see Chapter 4.3.4 for more information about hollow fine fiber membranes). Dupont also developed a flat-sheet form of the basic, linear aramid polymer used to make the B-9 hollow fine fiber membranes. This asymmetric membrane was called the B-15, and competed directly with CA membranes. DuPont withdrew their polyamide membranes from the market in 1991. DuPont sold their B15, A15, and developing X20 membranes to TriSep Corporation, who kept them for a few years before they licensed the FT30 chemistry from FilmTec (see Chapter 4.2.2.2).

4.2.2.2 Composite Polyamide Membranes

Composite membranes, sometimes called “thin film composite” or TFC® membranes are essentially a composite of two polymers cast upon a fabric support (TFC is a registered trademark of Koch Membrane Systems, Wilmington, MA). Today, cross-lined, fully aromatic polyamide membranes are the most popular RO membranes in use.¹¹ These membranes are generally formed using interfacial polymerization as shown in Figure 4.8 as developed by Cadotte (2, 19). A finely microporous substrate (typically polysulfone) is prepared using the Loeb-Sourirajan method described in Section 4.2.1.1, but without the annealing step such that the surface of the membrane does not “skin” over. The substrate is then exposed to monomers that are known to have high permeability to water but low permeability to salts, such as a polyamine. The amine is then immersed in a water-immiscible solvent containing a reactant, such as diacid chloride in hexane. The reactant and

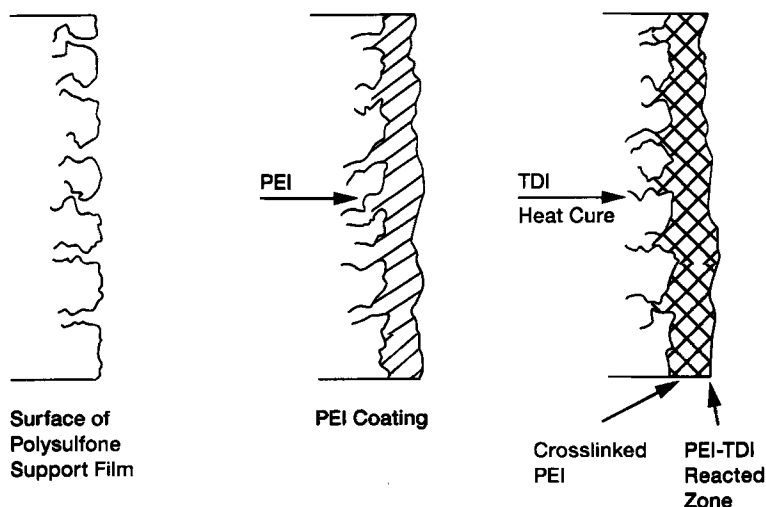


Figure 4.8 Interfacial polymerization, using polyethyleneimine (PEI) crosslinked with toluene diisocyanate (TDI) to form the NS-100 membrane as patented by Cadotte (U.S. Patent 4039440, August 2, 1977).

amine react at the interface of the water and organic solvent to form a highly-crosslinked thin film. This thin film formed on top of the substrate layer can be as thin as 400 to 1,000 Angstroms (0.04 to 0.1 microns). The substrate layer is typically 40 – 80 microns thick, while the overall thickness of the membrane, including the fabric backing, is about 1,500 to 2,000 microns thick.¹⁸ By comparison, a sheet of paper varies from about 1,000 to 2,750 microns, depending on the weight. Figure 4.9 shows a cross section of a polyamide, composite membrane.

Figure 4.10 shows the chemical structure of a polyamide membrane, namely the Dow Water and Process Solutions-FilmTec™ FT30 (FilmTec is a trademark of Dow Chemical Company, Midland, MI). This membrane is created using poly(phenylenediamine trimesamide) and trimesoyl chloride (U.S. patent 4277344, July 7, 1981). This chemistry is used in some form by virtually all major RO membrane producers.² Unlike the CA membrane, the polyamide membrane has negatively-charged functional groups. This serves to attract cationic polymers, and other cationic species, which can permanently foul the membrane. Unlike the linear polyamide produced by DuPont, the FT-30 membrane is highly cross-linked, as seen in Figure 4.10.

Table 4.2 lists the predominant characteristics of polyamide, composite membranes.

- The surface morphology of a polyamide membrane is rough, allowing for many areas where foulants can be captured and held by the membrane. See Figure 4.11.
- As discussed above, the charge on the polyamide membrane is negative, thereby attracting cationic polymer should it break through the pretreatment equipment.
- Unlike CA membranes, polyamide membranes cannot tolerate free chlorine or any other oxidizers. Some manufacturers quote 200 – 1,000 ppm-hrs of exposure until the membrane rejection is lost.²¹ This means after 200 – 1,000 hours of exposure to 1 ppm free chlorine, the membrane rejection will be unacceptably low. Chlorine attack is faster at alkaline pH than at neutral or acidic pH.
- Polyamide membranes can tolerate higher operating temperatures than CA membranes can. Polyamide membranes can be operated up to 45°F. However, the maximum temperature allowable is a function of pH. Table 4.3 lists maximum temperature as a function of pH. At temperatures greater than 45°C and pH greater than 11.5, the polyester support and polyamide layer themselves can be damaged.
- Polyamide membranes have a much broader pH range over which the membranes can be operated compared to CA membranes. Operational pH can range from 2 to 11, but may vary slightly for different membranes and different manufacturers.
- Operating pressure for polyamide membranes is generally lower than for CA membranes, ranging from about 150 to 400 psig (1 – 2.8 MPa). This is because the polyamide membrane has a thinner rejecting layer and its microporous support layer is extremely porous offering minimal resistance to permeate flow.
- Passage of salts and silica is much lower for polyamide membranes than for CA membranes. Some membranes can achieve 99.7+% rejection on a standard test solution (2,000 ppm NaCl solution at 225 psi

(1.6 MPa), 77°F (25°C), pH = 8, and 15% recovery). The salt passage at this rejection is only 0.3%, while the salt passage for high-rejection CA membranes at 98% rejection is 2%. So, the salt passage is nearly 7-fold lower for polyamide membranes than for CA membranes. See Chapter 3, Table 3.2 for complete rejection capabilities for polyamide membranes.

- An example of a polyamide membrane is the FilmTec FT-30 membrane.

4.2.3 Improvements to Polyamide, Composite Membranes

There have been several improvements made to polyamide, composite membranes that have enhanced their performance. Perhaps the most

Table 4.2 Characteristics of polyamide composite RO membranes.

Property	Value for PA Membranes
Membrane Type	Homogenous asymmetric, thin-film composite
Salt Rejection (%)	~98+
Silica Rejection (%)	~96+
pH Range	2–12*
Feed Pressure (brackish membrane)	145 – 400 psi
Temperature Tolerance	Up to 45°C**
Surface Charge	Negative (anionic)
Chlorine Tolerance	< 0.02 ppm
Biological growth	Causes membrane fouling
Fouling Tolerance	Fair
Surface Roughness	Rough

* Broader range possible for cleaning. Check with membrane manufacturer.

** Higher temperature possible for “heat sanitisable” membranes. Check with manufacturer.

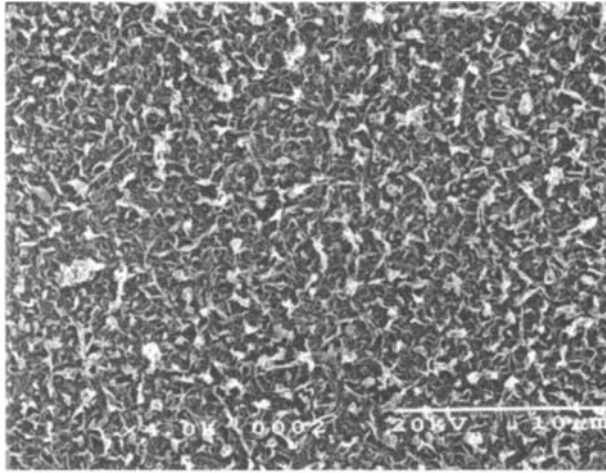


Figure 4.11 Rough surface morphology of polyamide composite RO membranes.

important improvement has come through advanced manufacturing techniques, which have allowed for thinner membranes with few imperfections. Thinner membranes exhibit higher flux rates at the same operating pressure than their thicker counterparts.

Some membrane manufacturers now offer “low fouling” membranes (see Chapter 4.4.2.3). These membranes exhibit better resistance to fouling with suspended solids. This is accomplished in several ways. Greater crosslinking of the polymer chain eliminates “hanging” (non cross-linked) function groups that can attract foulants.¹² Post-treatment of the membrane polymer, in a process sometimes called, “sizing,” is also used to minimize fouling of the membrane. Some manufacturers have created membranes with a lower surface charge and a smoother surface, both of which lead to minimal organic fouling.

Low-pressure membranes have also been developed. These membranes offer high flux at low temperatures and pressure albeit with some reduction in rejection (the permeability of polyamide membranes is a function of temperature, with lower water temperatures generally requiring higher operating pressures to maintain productivity—see Chapter 9.2). These low-pressure membranes allow for operation at low temperature at lower pressure than non low-pressure membranes.

4.2.4 Other Membrane Materials

Polyether urea (PEU) is another type of thin-film membrane. This membrane differs from polyamide membranes in the surface charge and morphology. Polyether urea membranes have a slightly positive charge to them. Further, the surface of a PEU membrane is smooth, similar to a CA membrane, thereby minimizing the potential for fouling. Hydranautics CPA membrane is an example of a polyamide/polyurea composite membrane.

4.3 Membrane Modules

Reverse osmosis membranes for industrial applications are typically modularized using configurations that pack a large amount of membrane area into a relatively small volume. This makes the RO system more economical to use in that the system requires a smaller footprint, and membranes can be replaced in smaller modules rather than system wide.

There are four basic forms for RO membrane modules: Plate and frame, tubular, spiral wound, and hollow fine fiber. These four configurations are summarized in Table 4.3 and discussed below. Additionally, some manufacturers have developed other module configurations that are briefly discussed in Chapter 4.3.5.

Table 4.3 Brief comparison of four basic RO membrane module configurations.²²

Property	Plate-and-Frame	Tubular	Spiral Wound	Hollow Fine Fiber
Packing Density, ft ² /ft ³	45–150	6–120	150–380	150–1,500
Potential for Fouling	Moderate	Low	High	Very High
Ease of Cleaning	Good	Excellent	Poor	Poor
Relative Manufacturing Cost	High	High	Moderate	Low

4.3.1 Plate and Frame Modules

Plate and frame RO modules are typically used for specialty, high suspended solids applications and are not generally found in water-purification facilities. These modules consist of flat sheets of membrane that are modularized into plates, typically two membranes placed back to back per plate. The plates are then stacked within a framework for support. There are patterned spacers materials that are used to keep the membranes from sticking to each other and providing open channels for the feed and product water to flow through. Figure 4.12 shows a typical plate-and-frame membrane module.

Characteristics of plate and frame modules are discussed below.

- These membrane modules are expensive per unit membrane area. This is because of a lot of hardware is used for relatively little membrane area.
- They are relatively easy to clean, hence their use in high suspended solids applications. Cleaning in-situ is possible but does not offer the best removal of foulants and scale. The best membrane cleaning involves removing the plates from the frame and hand-cleaning each individual flat sheet of membrane.

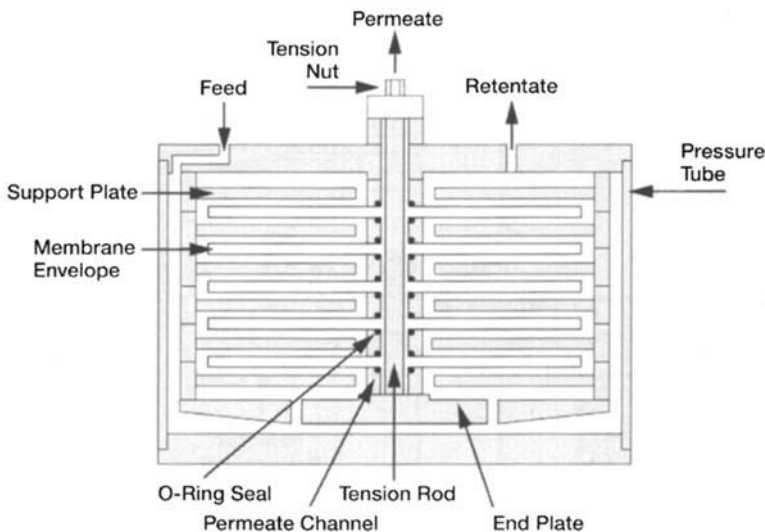


Figure 4.12 Plate-and-frame membrane module. *Courtesy of Elsevier.*

- These modules tend to foul because of the “dead” areas within the modules where cross-flow is not high enough to scour the surface free of debris.
- These and tubular modules offer the low packing density (membrane area per unit volume). Typical packing density is less than about $45 - 150 \text{ ft}^2/\text{ft}^3$ for plate-and-frame modules.²²

4.3.2 Tubular Modules

Tubular modules are also used for specialty, high-solids applications typically found in the food and biological processing industries. Tubular modules range from $\frac{1}{2}$ - to 1-inch (1.3 – 2.6 cm) in diameter with the membrane feed side on the inside of the tube. Packing densities run about $6 - 120 \text{ ft}^2/\text{ft}^3$.²²

Figure 4.13 shows how a tubular module is assembled.²³ These modules essentially resemble a shell-and-tube heat exchanger, with the RO feed water on the tube side and RO permeate on the shell side. The membrane tubes are supported by perforated stainless steel tubes through which the permeate exits.

Characteristics of tubular membranes are described below.

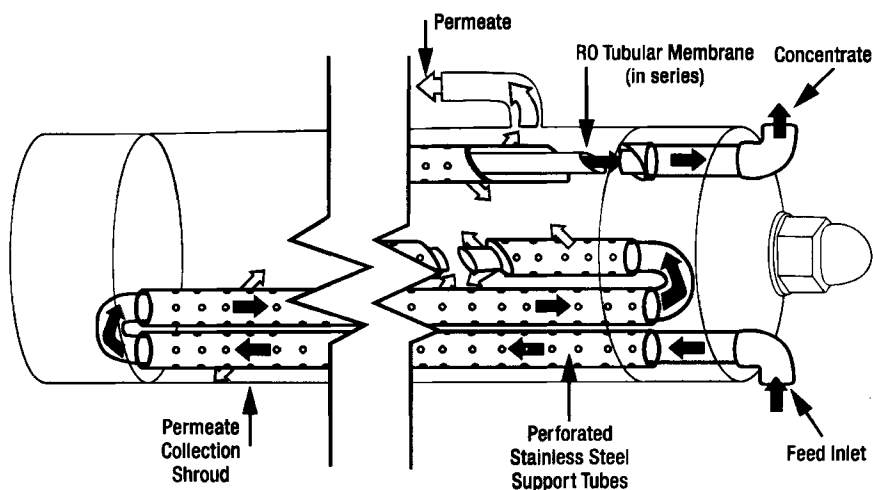


Figure 4.13 Tubular RO membrane module. Membrane tubes are placed in series in the housing.

- These modules are relatively expensive per unit membrane area. Again, the amount of hardware used per membrane unit area is significant.
- These modules are easy to clean. Typically, a sponge ball is shot down through the feed channel or tube to physically remove debris from the surface of the membrane. In most tubular applications, the membranes need to be cleaned on a daily basis. This is because the nature of the solution being treated by the membranes generally contains high concentrations of suspended solids and organics, which collect on the membrane. High feed flow rates, up to 20 gpm per tube, are necessary to achieve high cross-flow velocity and minimize fouling.¹¹

Most tubular membrane modules are used for specialty microfiltration (MF) and ultrafiltration (UF) applications rather than RO due to the lower packing density of this type of module and because MF and UF typically treat higher-solids feed water (see Chapter 16.1).

4.3.3 Spiral Wound Modules

Spiral wound membrane modules are the most common type of module used for RO today. The major advantage of a spiral wound module is that the packing density is fairly high, about 150 – 380 ft²/ft³,

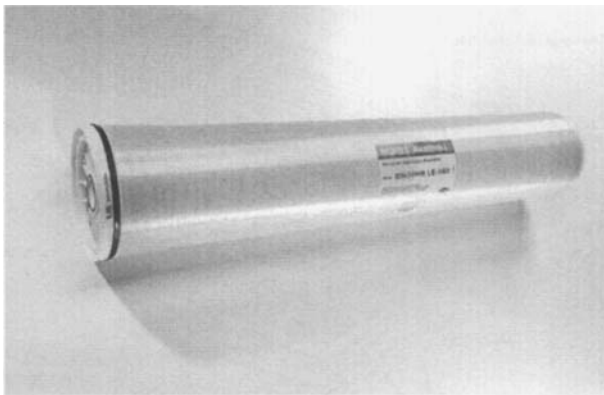


Figure 4.14 Eight-inch diameter spiral wound membrane module. *Courtesy of Dow Water and Process Solutions.*

higher than for plate and frame or tubular modules.²¹ Figure 4.14 shows an 8-inch diameter spiral wound membrane module.

Figure 4.15 shows the cross section of the spiral wound module.²⁴ The spiral construction starts with two sheets of membrane placed back to back with a nylon tricot mesh spacer material in between. This tricot spacer provides the permeate channel for the membranes. These sheets of membrane and spacer are glued on 3 sides so that the permeate can only exit the spacer on one side. This set of membranes and spacer is called a “leaf.” Leaves are then placed together with a low density polypropylene mesh spacer to provide the feed/reject channel for the membranes. The thickness of the mesh feed spacer can be adjusted from 28 mils to 34 mils to accommodate higher solids influent water (thicker feed spacers are more forgiving with respect to fouling with suspended solids than thinner spacers—see Chapter 4.4.2.3). The entire collection of leaves and mesh feed spacers are then wrapped around a perforated permeate collection tube so that the open side of the leaf is toward the perforated permeate tube (see Figure 4.16). Note that an 8-inch diameter membrane module has about 16 leaves, and each leaf is about 50 inches in length.

Influent that enters the spiral wound module does so tangentially to the membrane surface and the reject exits the module at the end opposite of the influent. Water that permeates the membrane does so perpendicularly to the surface of the membrane and is collected

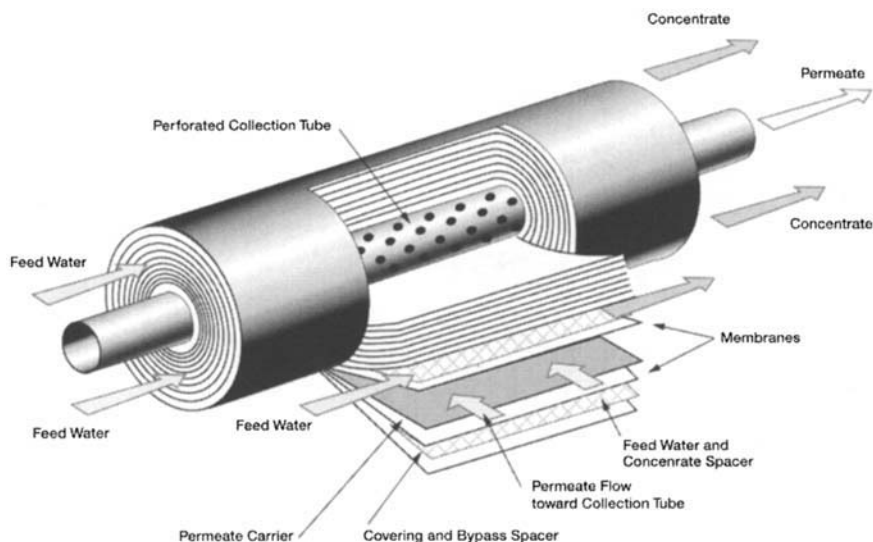


Figure 4.15 Deconstructed spiral-wound RO membrane module.

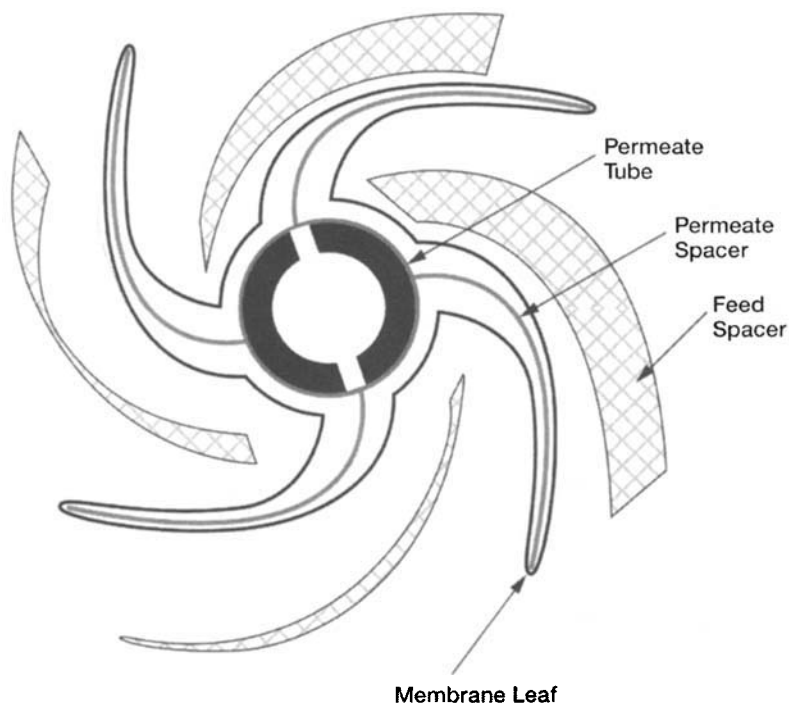


Figure 4.16 Spiral-wound RO membrane module showing leaves before winding.

in the permeate spacer and then spirals into the perforated permeate tube. Permeate usually exits the module through only 1 end of the permeate tube (this simplifies piping).

Not shown in Figure 4.15 are the end caps to the membrane modules. End caps are placed on each end of a membrane module. There are various forms of end caps, also called anti-telescoping devices or ATDs. The purpose of the end caps or ATDs is to prevent the membranes and spacers from telescoping under high differential pressure drop (see Chapters 11.3.1.3 and 12.3). Telescoping occurs when the membranes and spacer devices slide past each other and form a cone-shaped end rather than a flush end (see Chapter 11.3.1.3 and Figures 4.17a and 4.17b for more information about telescoping). The result is often damaged membranes that leak feed water into the permeate. Despite the ATDs, if the pressure drop is high enough, telescoping can occur. Telescoping can be uniform, as shown in Figure 4.17a or it can involve protruding spacers (Figure 4.17b) and membranes (Figure 4.17c) (see also Figure 14.11a). Figure 4.18a and 4.18b show two different styles standard of ATDs.

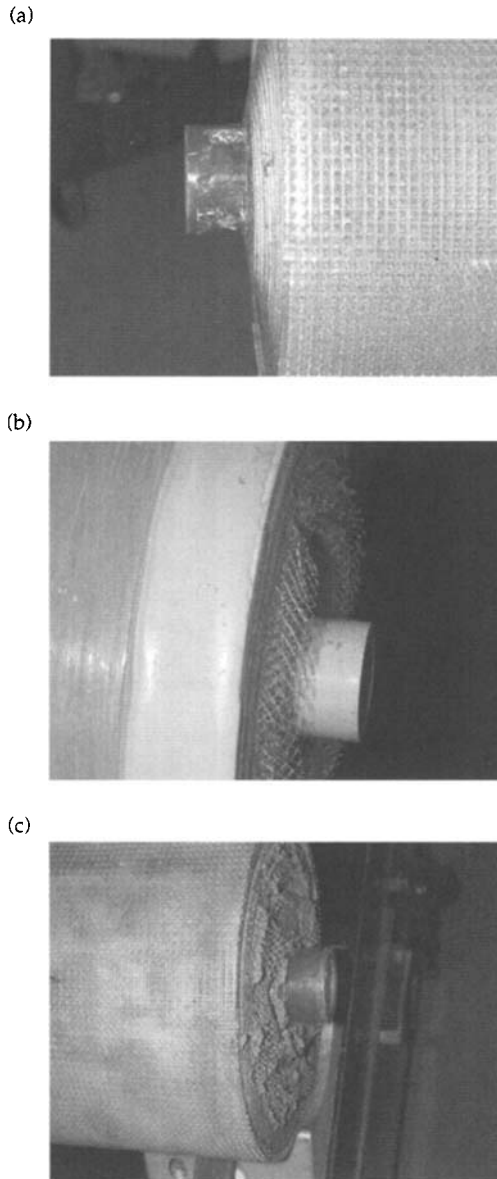
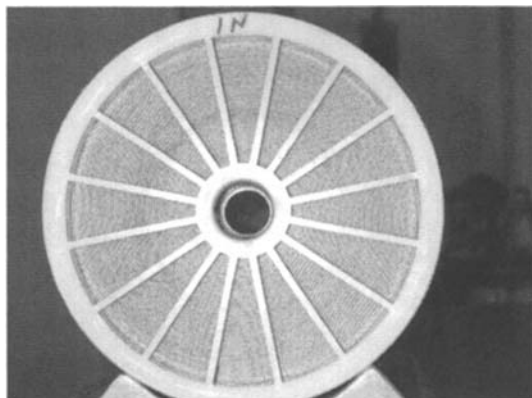


Figure 4.17 Uniform telescoping (a), protruding feed spacers (b), and protruding spacers and membranes (c).

Modules are connected to each other using an interconnector adaptor, as shown in Figure 4.19. The interconnector has O-rings at either end to ensure a tight seal with the module ATDs. These O-rings can roll upon installation into membrane modules, thereby

(a)



(b)



Figure 4.18 Two styles of standard anti-telescoping devices (ATDs).

allowing feed water to mingle with permeate. Great care should be exercised when loading membranes to prevent rolling the O-rings. Lubrication can some times be helpful to minimize friction and rolling (see Chapter 6.3).

Figure 4.20 shows the ATDs from a new FilmTec *iLEC*[™] (Interlocking End Cap) membrane module (*iLEC* is a trademark of The Dow Chemical Company, Midland, MI). The *iLEC* modules join ATDs by twisting them together directly rather than relying on interconnector adaptors, the way standard modules are connected together. As shown in the figure, the ATD *iLEC* end cap has an integral O-ring that cannot be rolled or pinched during installation. Furthermore,

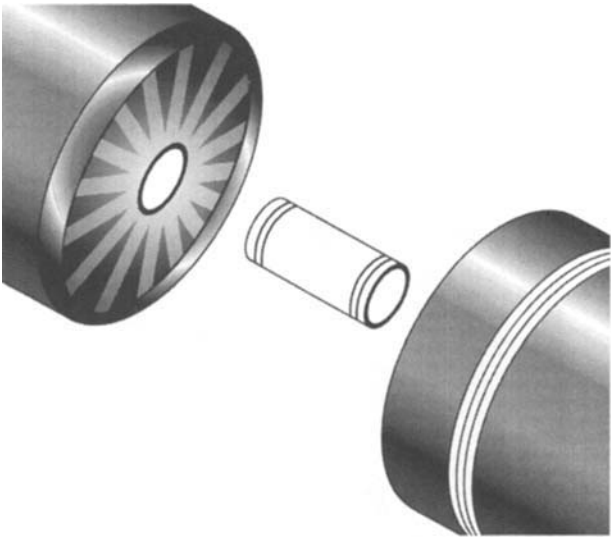
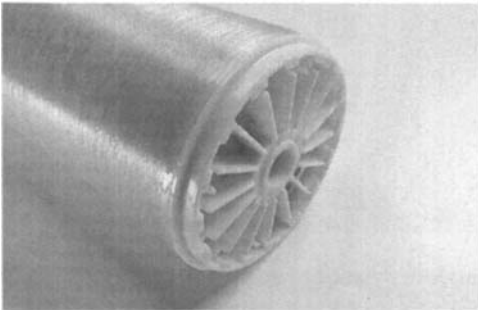


Figure 4.19 Placement of module interconnector adaptor for standard ATD end caps.

(a)



(b)

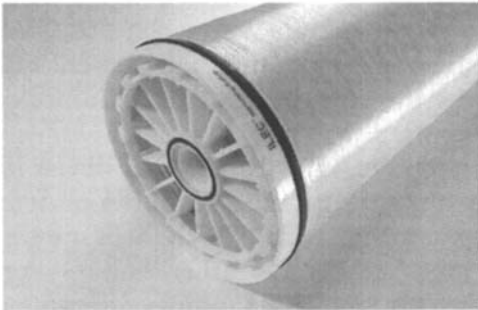


Figure 4.20 Dow Water Solutions-FilmTec *iLEC* ATDs with integral O-ring.
Courtesy of Dow Water and Process Solutions.



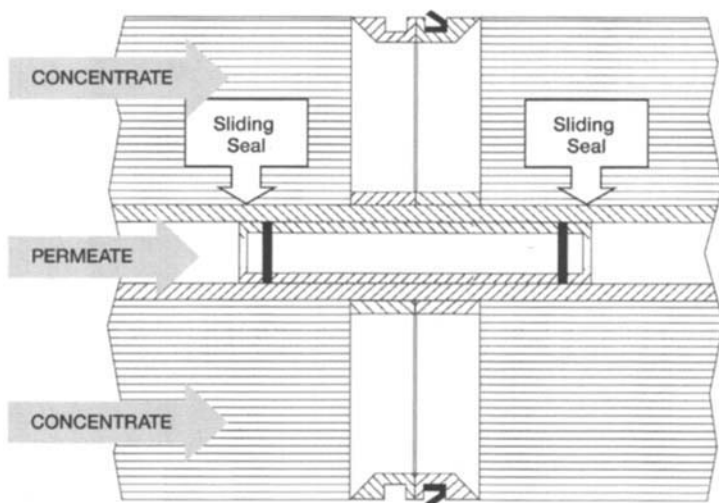
Figure 4.21 Use of strap wrench with *iLEC* membranes. *Courtesy of Crossbow Water Technologies.*

water hammer cannot wear on the O-ring as it does on the O-rings on conventional interconnectors, which leads to fewer leaks of feed water into the permeate (see Chapter 6.2). In fact, Dow Water Solutions' RO design program, ROSA, projects higher-quality permeate from an *iLEC* membrane than from the same membrane material in a non-*iLEC* configuration.²⁵

A strap wrench is used to hold one *iLEC* module in place as the other *iLEC* module is twisted either onto or off of the first module as illustrated in Figure 4.21.

Figure 4.22 shows the flow characteristics for standard ATDs and the *iLEC* ATD. The reduced diameter of internal couplers and vessel adapters for the standard ATDs accounts for more than 70 percent of the permeate-tube pressure drop in some systems.²⁶ The interlocking *iLEC* ATD design eliminates these restrictions, imposing less permeate backpressure, resulting in lower operating pressure requirements.

(a)



(b)

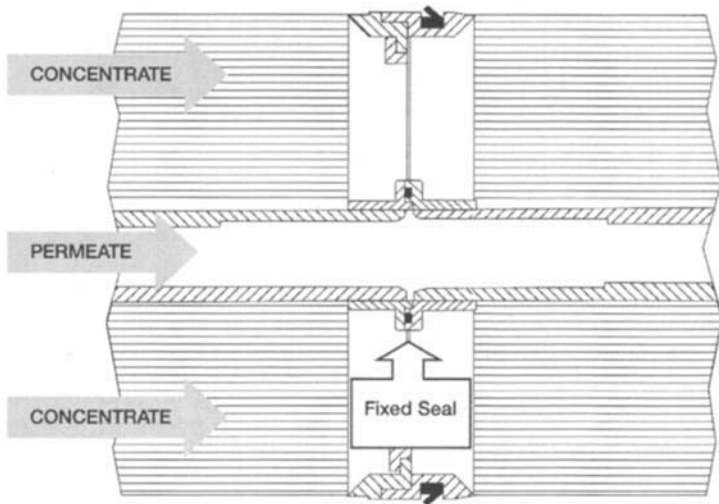


Figure 4.22 Flow characteristics through standard ADTs and iLEC ATDs. *Courtesy of Dow Water and Process Solutions.*

All spiral wound modules also have a brine seal (see Figures 4.23a and 4.23b). The brine seal is a U-cup-shaped rubber gasket material that is used to prevent feed water from passing by on the outside of the membrane module, thereby preventing feed water from bypassing the membranes (see Figure 4.24). The brine seal

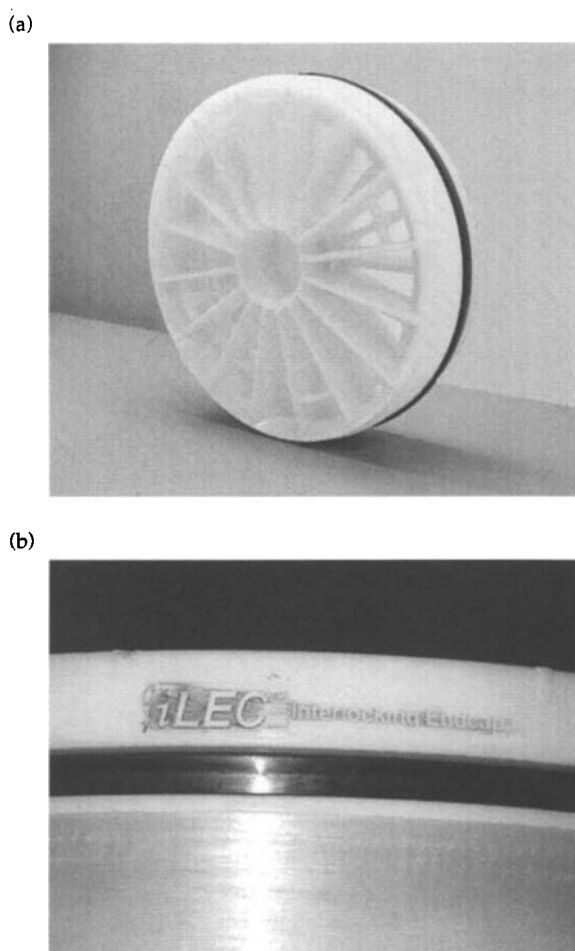


Figure 4.23 Two views of the “U cup” brine seal.

is located at the inlet end of the membrane module with the “U” facing the oncoming feed water.

The standard spiral wound module is 8 inches in diameter with a 40-inch length. There are also 4-inch and 18-inch diameter industrial membrane modules available (2.5-inch diameter modules are available for tap water or home-use applications). Koch Membrane Systems also makes a 60-inch long, 8-inch diameter module called a Magnum® and an 18-inch diameter by 60-inch length module called a MegaMagnum® (Magnum and MegaMagnum are trademarks of Koch Membrane Systems, Inc., Wilmington, MA). Figure 4.25

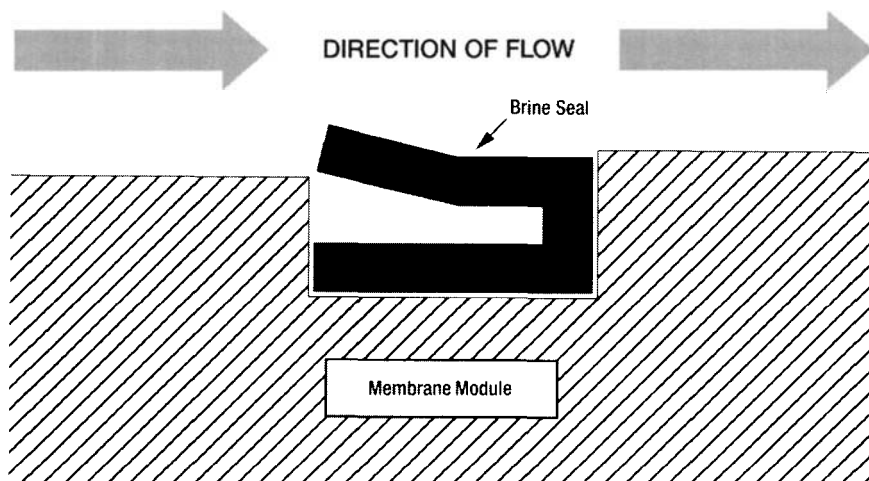


Figure 4.24 "U-cup" brine seal.

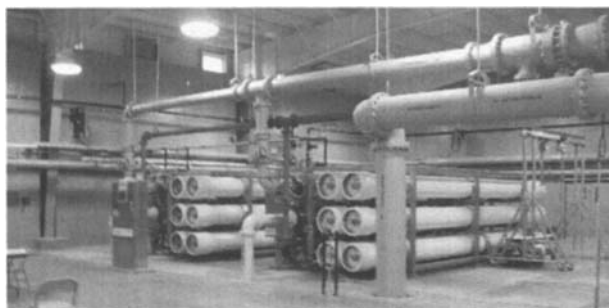


Figure 4.25 Two-train MegaMagnum RO system. Each train is capable of providing 1,390 gpm product flow when a 28-mil spacer is used, and 1,340 gpm when a 31-mil spacer is used. *Courtesy of Koch Membrane Systems.*

shows two trains of MegaMagnum modules. Each train is capable of providing 1,390 gpm product flow when a 28-mil spacer is used, and 1,340 gpm when a 31-mil spacer is used (see Chapter 4.4.2.3 for details about different thicknesses of feed spacers).

Spiral wound modules are typically covered in fiberglass to protect the leaves (exceptions being sanitary modules, see Chapter 4.4.2.6). Because of the materials of construction (namely the adhesives used) and the potential for "annealing" the membrane, the maximum operating water temperature is limited to 45°C.²⁷

Characteristics of spiral wound modules are described below.

- Spiral wound modules are moderately expensive due to the complexity and engineering involved. However, cost per unit membrane area is relatively low.
- Spiral wound modules can be difficult to clean. There are dead spaces within the module where high velocity cannot scour the surface of the membrane, and cleaning solution does not mix well to remove debris.
- Automated manufacturing of the membrane modules has allowed for more membrane area per unit volume and for higher-quality modules. This is because automation allows for more precise glue line application on the membrane leaves. A typical industrial module that is 8-inches in diameter and 40-inches long can hold up to 440 ft² of membrane area when automated manufacturing is employed (see Chapter 4.4.2.5).

Spiral wound modules are not themselves pressure vessels and therefore are placed in an external pressure vessel or “housing” for use. These pressure vessels are rated for the duty they will operate under, be it water softening, brackish water RO, or seawater RO (see Chapter 16.2 for membrane softening or “nanofiltration”). See Chapter 6.3 for a more detailed discussion about pressure vessels. Figure 4.26 shows a cut-away of a pressure vessel with a cut-away spiral-wound membrane module inside. Figure 4.27 shows the end of a spiral wound module in the pressure vessel without the pressure vessel end caps. Figures 4.28a and b show the pressure vessel end caps in place, one side with permeate effluent piping and one without this piping.

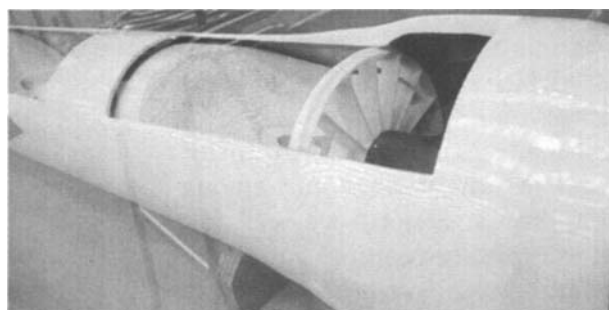


Figure 4.26 Cut-away of a pressure vessel with a spiral-wound modules inside.

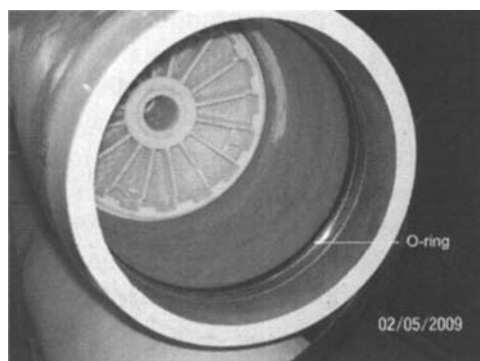


Figure 4.27 Spiral wound module in pressure vessel without pressure vessel end caps installed. The O-ring is used to seat the end cap in place. *Courtesy of Nalco-Crossbow LLC.*

4.3.4 Hollow Fine Fiber Membrane Modules

Hollow fine fiber RO modules are membranes formed into very small-diameter tubes, with an outside diameter of about 85 microns and an inside diameter of about 42 microns.¹⁵ The fibers resemble human hair and can be as flexible. The membrane “skin” or thin film is on the outside of the fiber. This skin is about 0.1 to 1 micron thick.¹⁵ Figure 4.29 shows a cross section of such a fiber.

Figure 4.30 shows a hollow fine fiber membrane module. The fibers are folded in half and the open end of each fiber is “potted” in epoxy “tube sheet,” while the folded end is potted in an epoxy, non-porous block. Feed to the module is outside in, which requires less strength on the part of the fiber than inside-out flow would. Also, the pressure drop on the outside of the fibers is much less than would be in the inside of the fiber (which is known as the lumen).

Characteristics of hollow fine fiber modules are described below.

- These modules are relatively inexpensive due to the high surface area per unit volume achievable with this configuration.
- These modules are relatively difficult to clean. There are several “dead” areas in and among the hollow fibers where flow does not reach well. Hence, higher-quality feed is typically required for these modules, such as seawater or brackish well water rather than high solids surface water.
- Packing density is extremely high, on the order of 150 – 1500 ft²/ft³.²²

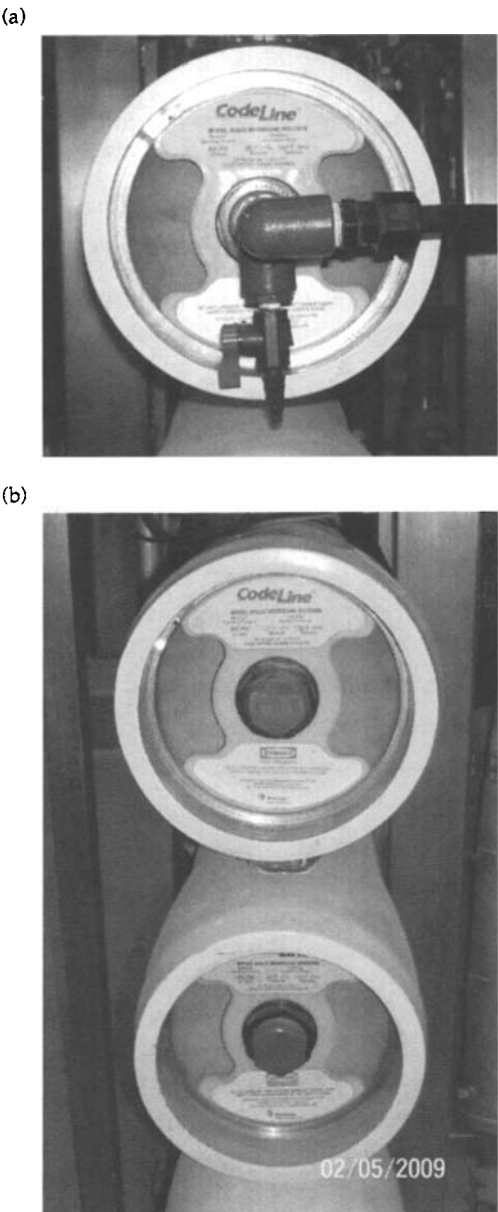


Figure 4.28 Pressure vessel end caps installed a) with permeate effluent piping and b) without permeate effluent piping (permeate exits the pressure vessel at one end only). *Courtesy of Nalco-Crossbow LLC.*

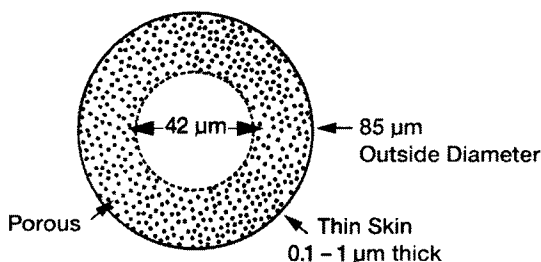


Figure 4.29 Cross section of a hollow fine fiber RO membrane.

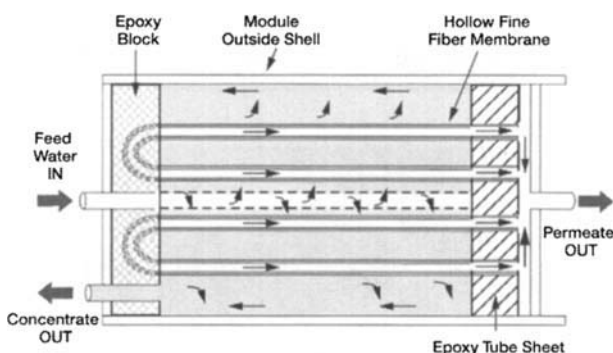


Figure 4.30 Simplified cross section of a hollow fine fiber RO membrane module.

As discussed in Chapter 4.2.2, DuPont introduced linear aromatic polyamide membranes in hollow fine fiber form as the B-9 (brackish water) and B-10 (seawater) Permeators. These Permeators were available in 4-, 8- and 10-inch diameter models. The 4-, 8-, and 10-inch B-9 Permeators were capable of producing 4,200, 16,000, and 25,000 gallon per day of permeate, respectively, at 75% recovery (standard test conditions: 1,500 ppm NaCl at 400 psig and 25°C).²⁸ Permeators ranged from about 47 inches to 53 inches in length. DuPont discontinued these modules in 2001.

Currently, Toyobo markets the Hollosep® cellulose triacetate hollow fine fiber for RO applications (Hollosep is a registered trademark of Toyobo Company, Ltd, Osaka, Japan).

4.3.5 Other Module Configurations

Some manufacturers have developed unique module configurations that rely on novel methods of introducing turbulence into the feed

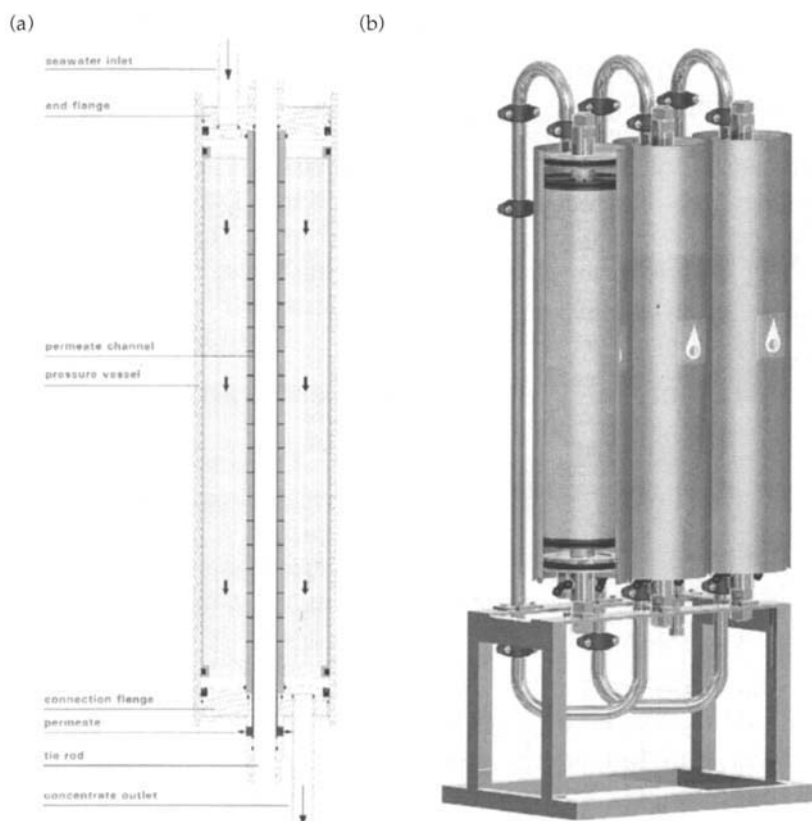


Figure 4.31 ROCHEM ST module, a modified spiral wound module, showing a) a cross-section and b) the module within a pressure vessel housing. *Courtesy ROCHEM Ultrafiltrations Systeme GmbH.*

stream as a method of minimizing concentration polarization. These configurations are generally suited to treat more difficult waters, such as waters containing high concentrations of suspended solids.

ROCHEM RO-Wasserbehandlung GmbH (ROCHEM) has developed three module configurations which offer reduced rates of membrane fouling. This is achieved using open feed flow channels and/or short feed flow water paths followed by a 180-degree flow reversal that introduces turbulence. Figure 4.31a is a diagram of the ROCHEM ST module, which features an open feed channel that minimizes fouling and pretreatment requirements as well as allows for easy membrane cleaning. These modules are used to treat waters ranging from brackish and seawater to land-fill leachate. Figure 4.31b shows the ROCHEM module in its pressure vessel housing.

New Logic Research, Inc. has developed a vibrating membrane disk module using what is called the Vibratory Shear Enhanced Processing (VSEP) technique to minimizing depositions of suspended solids on the membrane surface. The VSEP membrane stack consists of parallel membrane disk “leaves” separated by gaskets. The stack vibrates at 53 Hz and the amplitude of oscillation is $\frac{3}{4}$ to 1-1/4 inches thereby introducing turbulence into the pressurized feed. The VSEP system is used for a variety of applications including boiler feed water, RO reject, latex concentration, and acid mine drainage.

4.4 Commercially-Available Membranes

Several manufacturers currently supply RO membranes in the United States. Table 4.4 provides a brief description of several current US industrial RO membrane manufacturers.

Many varieties of spiral-wound, polyamide-composite membranes are available to suit different feed water conditions. Membranes discussed here include:

- Seawater membranes
- Brackish water membranes
- Brackish, low-energy membranes
- Brackish, low-differential pressure membranes
- Brackish, low-fouling membranes.

4.4.1 Seawater Membranes

Seawater membranes are used to treat high-salinity (35,000 to 50,000 ppm total dissolved solids (TDS)) feed waters. These membranes can operate at pressures up to 1,500 psi. Typical membrane test conditions are as follows:

- Feed water concentration: 32,000 ppm NaCl (sodium chloride)
- Operating pressure: 800 psi
- Temperature: 77°F
- Feed water pH: 6.5 – 8
- Recovery per module: 8 – 10%

Test conditions are important to take note of as these are the conditions under which rated performance is based. Operating under

Table 4.4 Current US manufacturers of industrial RO membranes.

Manufacturer	Location	Brand	Membrane Material	Spiral Wound Configuration
Koch Membrane Systems, Inc.	Wilmington, MA, San Diego, CA	<ul style="list-style-type: none"> • Koch • Fluid Systems 	PA CA	Standard Sanitary
Toray Membrane USA	Poway, CA		PA	Standard Sanitary
Dow Water & Process Solutions	Edina, MN	<ul style="list-style-type: none"> • FilmTec 	PA	Standard Sanitary
Hydranautics	Oceanside, CA		PA	Standard Sanitary
TriSep Corporation	Goleta, CA		PA CA LAPA	Standard Sanitary Custom
GE Water & Process Technologies	Vista CA, Minnetonka, MN	<ul style="list-style-type: none"> • Desal • Osmonics 	PA CA	Standard Sanitary
Sepro Membranes	Vacaville, CA		Proprietary	Standard*
Osmosis Technology, Inc.	Buena Park, CA	<ul style="list-style-type: none"> • Osmotik 	PA CA	Standard Custom

PA = Polyamide composite

CA= Cellulose acetate or cellulose triacetate

LAPA = Linear aromatic polyamide (developed by DuPont)

* also available in flat sheets

Table 4.5 Comparison of Koch Membrane Systems seawater membranes.²⁸

Koch Membrane Systems Seawater Membrane	Permeate Flow* gpd	Rejection* (%)
8040-HF-400	9,500	99.7
8040-SW-400	7,200	99.75

* for 400 ft² membrane module

different conditions will result in performance that differs from the rated performance. (Chapter 9 discusses the effect of varying operating conditions on the performance of RO membranes). Notice that there is not one uniform test condition to which all membrane manufacturers adhere. Therefore, because of the difference in pH and recovery under such membranes are tested, the rated performance of seawater membranes from different manufacturers cannot be directly compared.

Within the classification of seawater membranes, there are subsets of membrane that are rated for different performance. For example, Koch Membrane Systems offers a standard high rejection seawater membrane module plus a high-flow seawater element seawater membrane module. Table 4.5 lists the productivity and rejection for two seawater membrane types. Other membrane suppliers offer similar variety in seawater membranes.

4.4.2 Brackish Water Membranes

Brackish water membranes are designed to treat lower-salinity feed waters, up to about 4,000 to 5,000 ppm (TDS). Maximum operating pressure for brackish water membranes is typically 600 psi. These membranes are typically tested at the following conditions:

- Feed water concentration: 1,500 to 2,000 ppm NaCl (low energy membranes are tested at 500 – 2,000 ppm NaCl)
- Operating pressure: 225 psi (low energy membranes are tested at 100 – 150 psi)
- Temperature: 77°F
- Feed water pH: 6.5 – 7
- Recovery per module: 15%

As with seawater membranes, there is no one uniform test condition for all brackish water membranes of the same type. Thus, a direct comparison between manufacturers requires a close look at the test conditions.

Brackish water membranes also come in various types with different performance ratings. These specific types are discussed below.

4.4.2.1 Low-Energy Membranes

Low energy membranes are designed to reduce the energy required to generate permeate. In most cases, these membranes exhibit similar productivity but at lower operating pressures than standard RO membranes. Lower operating pressure is an advantage when energy costs are high or when the feed water temperature is low (lower water temperature reduces the water throughput if pressure is held constant, as discussed in Chapter 9.2). A limitation of the low-pressure membrane is that rejection is lower than the standard brackish water membrane. In some cases the rejection drops enough to double the salt passage as compared to standard brackish water membranes.

4.4.2.2 High-Rejection Membranes

High-rejection brackish water membranes offer several tenths of a percent higher rejection than standard brackish water membranes. While the standard rejection is typically about 99.0% to 99.5%, high rejection membranes can go as high as 99.7% rejection (some newer membranes now claim 99.75% rejection). Going from 99.5% rejection (0.5% salt passage) to 99.7% rejection (0.3% salt passage) decreases the salt passage by 67%. This can be critical in high-purity applications.

4.4.2.3 Low-Fouling Membranes

Low-fouling membranes are available from some manufacturers. These membranes can be modified in several ways to reduce the potential for fouling them with contaminants in the feed water. Chapter 4.2.3 describes these modifications in more detail.

An example of a low-fouling membrane is the Hydranautics low-fouling composite LFC3-LD membrane (see Chapter 4.4.2.3 for a description of the LD (low-differential-pressure) membrane). This membrane exhibits the same throughput as the standard

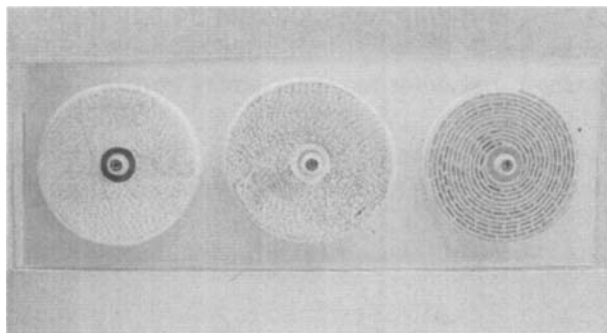


Figure 4.32 Cross section of 4-inch diameter membrane modules with feed spacers of various thicknesses, ranging from 28-mils, to 31-mils, to 34 mils-thick when viewed left to right.

high-rejection Hydranautics CPA3 membrane and has slightly higher rejection.³⁰ The difference is in the surface charge of the membrane. The standard brackish water membrane has a negative charge while the low-fouling membrane has a neutral surface charge. This will minimize fouling with cationic polymers and surfactants as well as other positively-charged species that will foul a negatively-charged membrane. Additionally, the membrane module is constructed using 31-mil feed spacers rather than the standard 28-mil spacer. Thicker feed spacers are more forgiving to fouling with suspended solids than thinner spacers. Figure 4.32 shows the cross section of 4-inch diameter membrane modules with feed spacers of various thicknesses. It is easy to see how the thinner the feed spacer, the more prone to fouling the corresponding module will be.

4.4.2.4 Low-Differential-Pressure Membrane Modules

Low-differential-pressure membrane modules can be considered a subset of low-fouling membranes. These low-differential-pressure membrane modules typically have a thicker feed spacer. Instead of the standard 28-mil thick spacer, these low-differential-pressure membranes have 31- or 34-mil thick spacers. There is less resistance to flow through the feed channels, resulting in lower pressure drops through the membrane modules. Furthermore, the feed channels will not plug as quickly with suspended solids, foulants, or scale. Examples of low-differential-pressure membrane modules are the FilmTec BW30-400-34i (with a 34-mil feed spacer) and the Hydranautics CPA3-LD (with a 31-mil feed spacer).

4.4.2.5 *High-Productivity Membrane Modules*

High-productivity membrane modules contain more membrane area than standard brackish water membranes despite fitting into the same size membrane module. Higher membrane area is achieved using more sophisticated module-assembly techniques. Careful positioning of the glue lines on the membrane leaves (see Chapter 4.3.3) and automated module assembly are two improvements in module assembly that have allowed for the inclusion of more membrane area. While a standard brackish water membrane typically has about 365ft² of membrane area, high-productivity membrane modules may have 400ft² or up to 440ft² of membrane area. Productivity out of the membrane module is higher because of the additional membrane area. In general, a 400ft² membrane module produces about 10% more permeate than a 365 ft² membrane module produces under similar operating conditions.

High productivity can also be achieved with brief, measured exposure to free chlorine (see Chapter 8.2.1). Membrane manufacturers will sometimes treat their membranes with a very short exposure to free chlorine. This results in membranes that exhibit higher flux with no change in salt rejection. Longer exposure to free chlorine will result in a permanent loss of salt rejection. Note that exposure to free chlorine by the end user is a violation of the membrane warranty and should not be attempted to increase flux.

4.4.2.6 *Other Membrane/Module Types*

There are other types of membrane and modules available. These types spring from applications where the need is different than standard membranes can handle. Two common applications are boron rejection and sanitary processing.

- Boron rejection membranes: exhibit up to 90+% rejection of boron, while standard membranes reject about 50–70%.^(20, 29) These membranes are typically used for seawater applications where boron removal is a concern. Boron is difficult to remove with membranes because boron, which exists as boric acid, is not ionized at a typical seawater pH, 7.0 – 8.0, whereas the pK_a of boric acid is 9.14 – 9.25.²⁰
- Sanitary membrane modules: these modules have a net outer wrap rather than the standard fiberglass wrap and are sometimes referred to as “full fit” modules



Figure 4.33 Sanitary or “full fit” spiral wound membrane module. *Courtesy of Dow Water Solutions.*

(see Figure 4.33). The modules are designed with a bypass (no brine seal) to eliminate stagnant or “dead” areas within the module where bacteria can grow and foulants can accumulate; typically 20% of the feed flow by-passes the membrane modules.²⁵ Additionally, the modules are operated at higher vessel flows (higher cross-flow velocities) and higher differential pressures to keep them clean. The trade off is that these membranes are less efficient than conventional spiral-wound RO modules. Some sanitary membrane modules can be sanitized for short periods of time at temperatures up to 85°C (recall that the maximum temperature for a standard spiral wound module is 45°C). This is because of changes in the materials of construction, including the permeate spacer. Standard permeate spacers soften as they are heated; those used for sanitary applications do not. Note that the RO membrane materials are the same and they anneal under heat, making them denser and more difficult to force water through them.³¹ Thus, sanitary membrane modules last about only about one to two years on average, whereas standard RO modules last up to three to five years in use. Sanitary applications include dairy, pharmaceutical, and biological processing.

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Basic Flow Patterns

Knowledge of the flow patterns for RO systems is fundamental to the understanding of how an RO system functions. Arrays, passes, recycle, and multiple trains are terms used to describe an RO system that are discussed in this chapter.

5.1 Arrays

Focusing on spiral wound membrane modules as the most common type of membrane modules used in industry today, an RO array or “skid” or “train” consists of a number of pressure vessels arranged in specific patterns. Figure 5.1 shows an array of 3 pressure vessels. The pressure vessels are arranged into 2 sets, with 2 pressure vessels in parallel followed by 1 single pressure vessel. The 2 sets of pressure vessels are in series. Each set of pressure vessels in parallel (even if there is only 1 vessel) is called a *STAGE*.

In theory, influent water to the RO system is split evenly among the pressure vessels in the first stage. Permeate from each pressure vessel in the first stage is combined and collected in a common header. The reject from the first stage becomes the influent to the second stage. Permeate from the pressure vessels in the second stage is collected and combined with permeate from the first stage to become the overall permeate from the system. The reject from the second stage becomes the reject for the entire system.

The RO system shown in Figure 5.1 is called a 2-stage array, or a 2:1 array, indicating that there are 2 stages (by the 2 numbers) and the first stage has 2 pressure vessels and the second stage has 1 pressure vessel. A 10:5 array would have 2 stages, the first stage would have 10 pressure vessels while the second stage would have 5 pressure vessels. A 4:2:1 array would have 3 stages, with 4 pressure vessels in the first stage, 2 pressure vessels in the second stage, and 1 pressure vessel in the third stage.

This type of array, the $(n):(n/2):(n/4)$ is called a “taper” or “christmas tree” configuration. The reason for the tapered design involves maintaining cross-flow velocity. The number of pressure

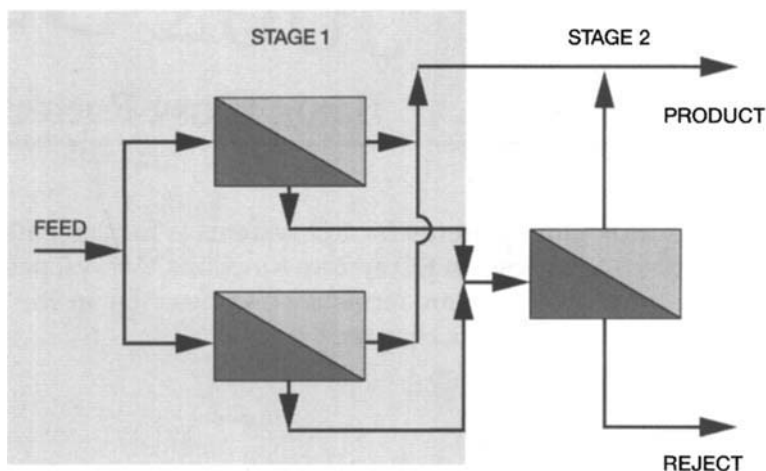


Figure 5.1 Two-by-one (2:1), 2-stage array with 2 pressure vessels in the first stage and 1 pressure vessel in the second stage.

vessels required for each stage is determined by the velocity or influent flow rate to that stage. To maintain good cross-flow velocity (Chapter 2.4), influent flow rates per pressure vessel need to be about 40–60 gpm, while the reject flow rate needs to be greater than about 16 gpm, for 8-inch diameter membrane modules (see Chapter 9.1 and Tables 9.2 and 9.3 for more detailed discussions). Thus, an influent flow of 100 gpm would require 2 pressure vessels in the first stage (see Figure 5.2). Ideally, the first stage recovers about 50% of the influent water (assumes six 8-inch diameter membrane modules in series), so that 50 gpm would be permeate from the first stage and 50 gpm would be reject. The reject is sent on to the second stage as its feed. Since the flow now is only 50 gpm to the second stage, only 1 pressure vessel is required. (If two pressure vessels were used, the feed flow rate per pressure vessel would drop to 25 gpm each, too low to maintain good feed cross-flow velocity.) The reject from the single pressure vessel in the second stage would be 25 gpm, well above the 16-gpm minimum concentrate flow rate per pressure vessel. Permeate from the second stage would be about 25 gpm, and adding that to the 50-gpm permeate from the first stage makes the overall recovery of the system 75 gpm or 75%.

Overall recovery from a 2-stage RO is typically about 75%. Higher recoveries (80%) can be reached provided the influent water is relatively free of suspended solids and scale formers. Recoveries higher than

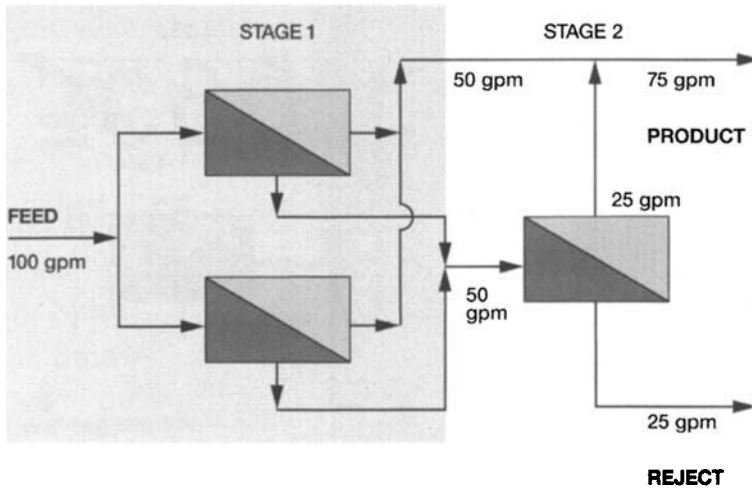


Figure 5.2 Approximate flow is distribution through a 2:1 array.

about 75-80% generally require more than 2 stages (again, assuming six, 8-inch diameter modules per pressure vessel).

Figure 5.3 shows how concentration changes through an RO system (assuming 50% recovery per stage, as in Figure 5.2). In this example, assume a feed concentration of 100 ppm total dissolved solids (TDS) and a membrane rejection of 98% TDS (2% salt passage). The permeate concentration out of the first stage would be 2% of 100 ppm or 2 ppm. To calculate the reject concentration, use the concentration factor. Recall from Table 3.1 concentration factor as a function of recovery. At 50% recovery after the first stage, the concentration factor is two. Thus, the reject concentration after the first stage is approximately 200 ppm. The feed to the second stage is then 200 ppm. The salt passage is still 2%, but now the concentration is double the original feed, so the permeate from the second stage is double the first stage permeate concentration, or 4 ppm. The recovery over the second stage is 50%, so the concentration factor is again two. This makes the overall reject concentration approximately 400 ppm. To calculate the permeate concentration, take 50 gpm permeate flow from the first pass and multiply by its concentration of 2 ppm. Add this to the result of multiplying the second stage flow of 25 gpm by the second stage concentration of 4 ppm. Divide by the total permeate flow of 75 gpm and the result is 2.67 ppm. Although the individual membrane rejection is 98%, the overall *system* rejection is 97.3%. This is because the 98% rejection refers to an individual membrane module under

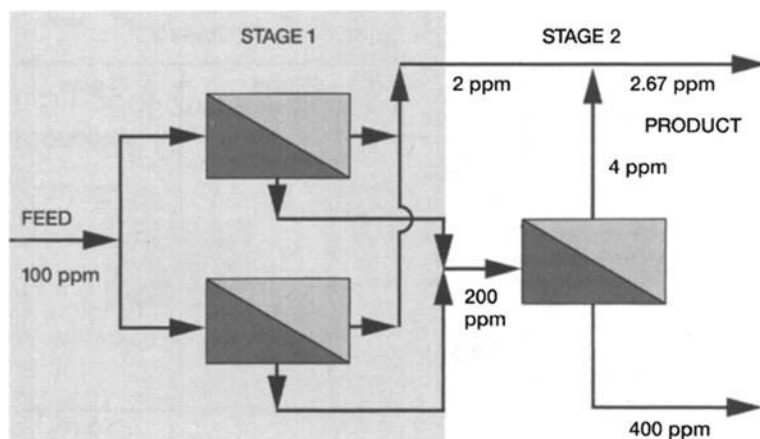


Figure 5.3 Concentration changes through an RO system assuming 98% rejection by the membranes.

test conditions (typically 10-15% recovery for brackish water membranes—see Chapter 4.4.2). Figure 5.4 shows in greater detail how the concentration changes per module through a single stage of an RO system. The data in Figure 5.4 assumes 11% recovery per module at 98% salt rejection.

In addition to observing how the flow and concentration change over a single RO stage, it is also interesting to see how the Langelier Saturation Index (LSI) changes with position in two-stage, 75%-recovery RO system (See Chapter 3.9 for a discussion about

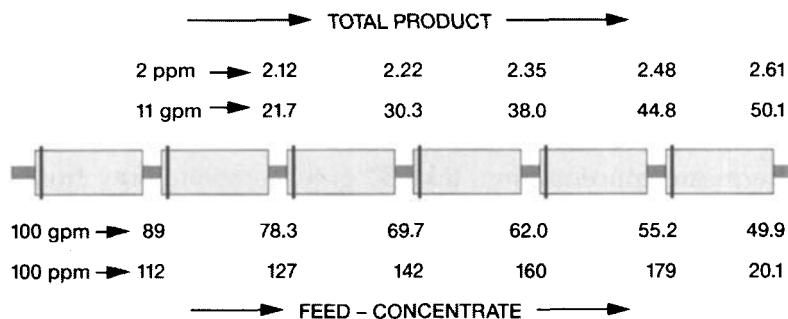


Figure 5.4 Module-by-module concentration and flow rate changes over a single RO stage. Assumes 11% recovery per module and 98% solute rejection.

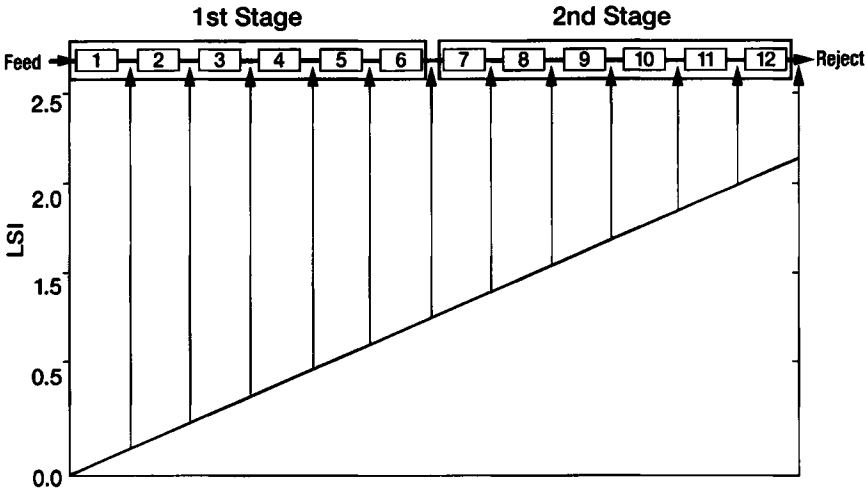


Figure 5.5 Langeleir Saturation Index (LSI) as a function of module position and recovery for a two-stage, 75%-recovery RO system. Assumes feed water conditions: 200 ppm calcium, 150 ppm bicarbonate, and pH = 7.0.

LSI). Figure 5.5 shows how concentrate LSI increases with increasing recovery through an RO system at 75% total recovery. The LSI increases linearly from zero with no recovery to over two at 75% recovery. Recall from Chapter 3.9 that if the LSI is greater than zero, the potential exists for scaling the membranes. Antiscalants, and in some cases acid feed, is required to minimize this potential (see Chapter 8.2.2).

A two- or three-stage RO system will usually remove about 96% to 98% of the total dissolved solids (TDS) in the water. The effluent water quality is generally high enough to send to low- to medium-pressure boilers without additional polishing. However, each application and feed water is different, and, therefore, water sampling and design projections should be conducted for every application to determine what the projected water quality will be (see Chapters 7 and 10).

5.2 Recycle

Figure 5.6 shows an RO array with concentrate recycle. A concentrate recycle is generally used in smaller RO systems, where the cross-flow velocity is not high enough to maintain good scouring of the membrane surface. The return of part of the concentrate to

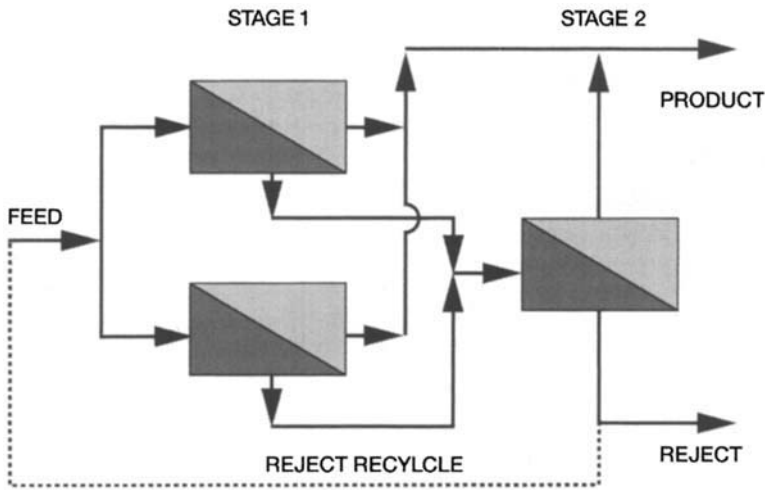


Figure 5.6 Two-by-one array with concentrate recycle.

the feed increases the cross-flow velocity and reduces individual module recovery, thereby reducing the risk of fouling.

Recycle has some disadvantages as well:

- Lower overall product quality. This is because relatively high-concentration reject is added to the lower-concentration influent.
- Larger feed-pump requirements, because the RO feed pump must now pressurize both the influent stream plus the recycled reject stream. As a result, the RO feed pump must be larger, which may mean higher capital for the RO system.
- Higher energy consumption, again because of the reject and influent streams coming together and must be repressurized. This results in higher operating costs for the system.

5.3 Double Pass

Double pass (or two-pass) refers to further purification of permeate from one RO by running it through another RO. The first RO, as described in Chapter 5.1, would be the first pass. Permeate from the first pass is then sent to another RO known as the second-pass RO.

The second-pass RO “polishes” the first-pass RO product to yield higher-quality water.

Figure 5.7 shows a double-pass RO system. The design principles for the second pass are generally the same as for the first pass. However, because of the low concentration of dissolved and suspended solids in the influent to the second pass, the influent and concentrate flows can be higher and lower, respectively, than for the first-pass RO system (see Chapters 9.4 and 9.5, and Tables 9.2 and 9.3). Because the reject from the second pass is relatively clean (better quality than the influent to the first pass), it is virtually always recycled to the front of the first pass. This minimizes the waste from the system and also improves feed water quality, as the influent to the first pass is “diluted” with the relatively high-quality second-pass reject.

Recovery of the second pass can be as high as 90% with only 2 stages. This high recovery can be achieved because of the relatively low-concentration of dissolved solids in the influent to the second pass. Overall system recovery will be 73% with 75% first pass and 90% second pass recoveries (recovery would be 67.5% without recycle).

A tank is typically required between the first and second pass systems. This is so that flows can equalize between the passes. However, if the number of first-pass skids is equal to the number of second-pass skids, a tank may not be required.

Some vendors place both passes on a single skid, thereby eliminating the RO feed pump to the second pass RO. The backpressure from the first pass is sufficient to provide the applied pressure required of the second pass. Care must be taken so that permeate backpressure does not exceed the applied influent pressure to the first pass, or osmosis rather than reverse osmosis will occur. Additionally, high back pressure can lead to delamination of the membranes (see Chapter 12.1.2.1 and Figure 12.1)

Inter-pass caustic injection is commonly used to drive out carbon dioxide from the first-pass RO permeate/second-pass feed. Carbon dioxide, a gas that is not rejected by the membranes, is converted to bicarbonate alkalinity, which is rejected (see Chapter 3.2). Removal of carbon dioxide is particularly important for applications that polish the second-pass RO permeate with ion exchange. The conversion and, therefore, elimination of carbon dioxide from the permeate will reduce the loading on the anion resin.

Effluent quality from a double-pass RO system is generally high enough to allow for direct use in 600 to 900 psi boilers. Higher pressure boilers (>1,000psi) and higher purity applications will

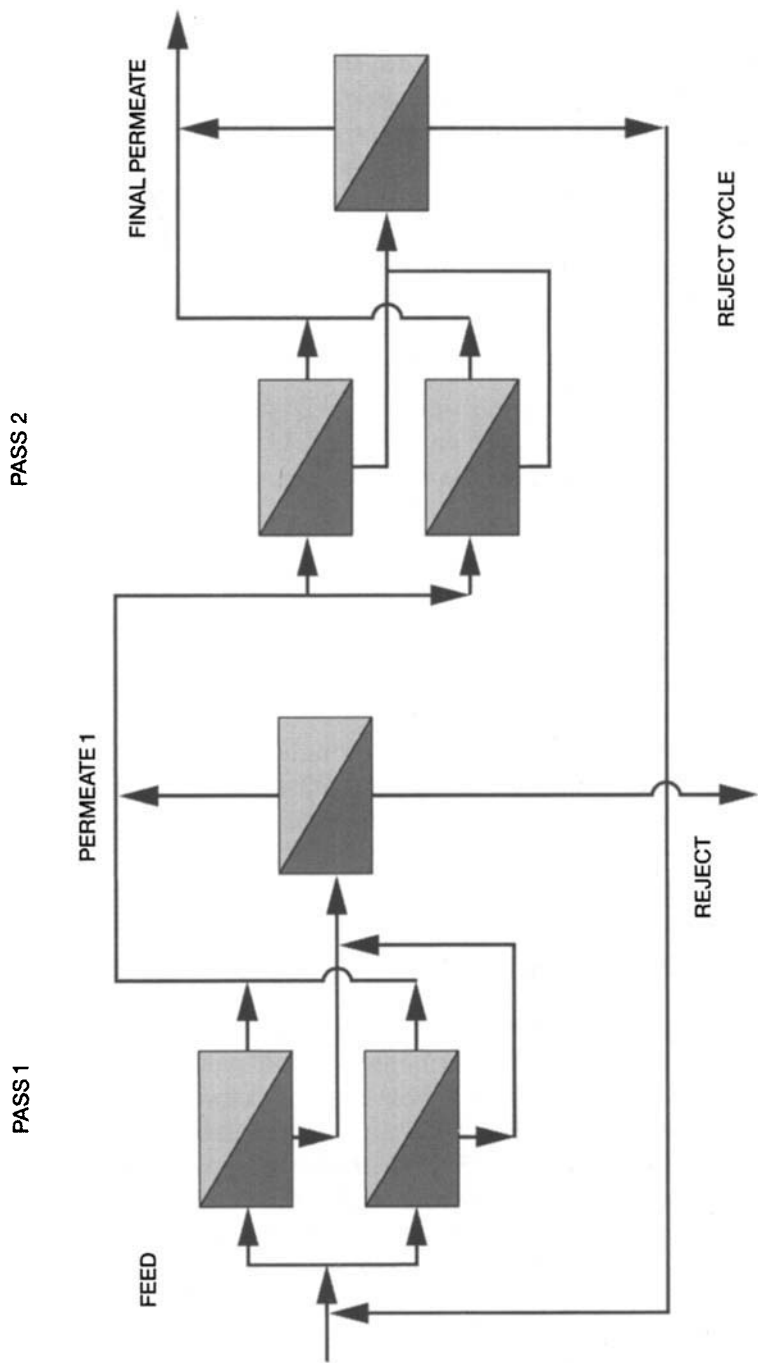


Figure 5.7 Double-pass RO.

still require some sort of post-treatment, typically a mixed-bed ion exchanger or electrodeionization (see Chapter 16.3).

5.4 Multiple Trains

Multiple trains or “skids” placed in parallel are used when larger flow rates need to be treated. For example, an 800-gpm RO might require 1 skid, if the vendor has that size skid in their inventory. Alternatively, one could use two 400-gpm RO skids to make up the 800 gpm production rate.

There is an advantage to using multiple skids in that multiple skids provide redundancy for the system; one skid can still be on line while the other is off line for cleaning or maintenance. Additionally, multiple skids can be used to juggle variable product water demands. See Chapter 15.2 for more information about variable demands and multiple RO skids.

The drawback to multiple skids is in capital and operating costs; the greater the number of skids, the higher the capital and operating/maintenance costs. These costs must be weighed against the ability to still provide water during shut down of any one skid for cleaning or maintenance. In some cases, multiple skids will make economic sense and in others, bringing in temporary equipment during shut down will make economic sense.

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Reverse Osmosis Skids*

An RO skid includes the pressure vessels in which the membrane modules are contained (see Chapters 4.3.3 and 6.3 for detailed discussions about pressure vessels). Skids also commonly include cartridge filters in a housing or housings and an RO feed pump, although combinations exist with just pressure vessels or pressure vessels with cartridge filters. Finally, there are included on the skid instrumentation and controls for the system. Figure 6.1 shows an RO skid with these components.

Figure 6.2 shows a detailed process flow diagram (PFD) for a 2:1 array RO system. The figure shows the major components of an RO system including instrumentation, control switches, and valves.

Components of an RO system discussed in this chapter include:

- Cartridge filters
- RO feed (booster) pumps



Figure 6.1 Reverse osmosis skid. *Courtesy of Nalco-Crossbow Water LLC.*

* With contributions by Mark Sadus
Nalco-Crossbow Water LLC, Glenwood, IL 60425 USA

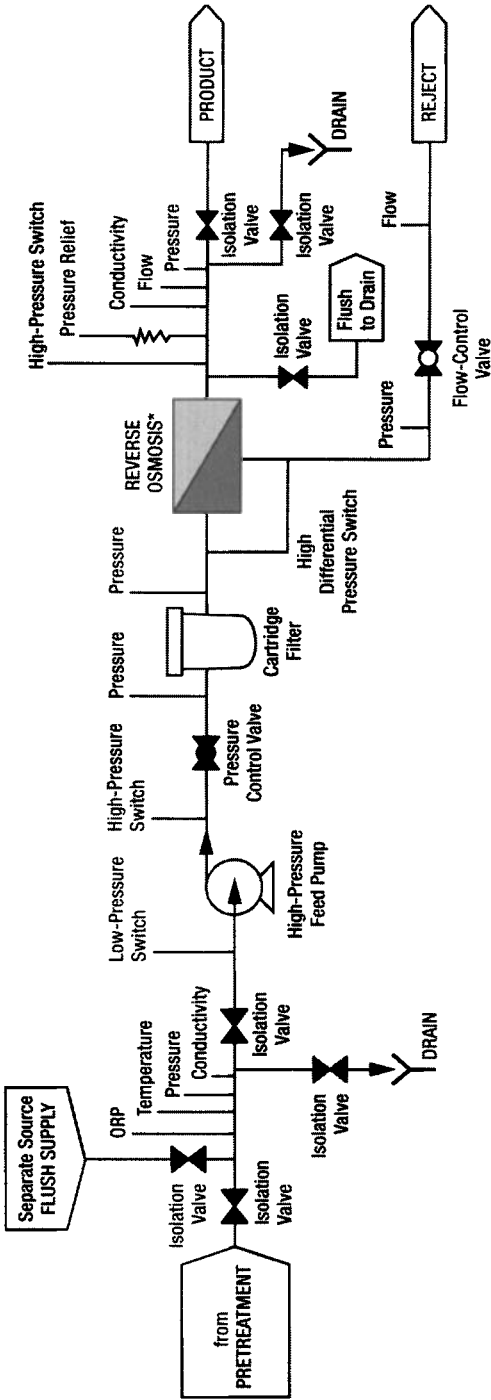


Figure 6.2 Sample RO process flow diagram (PFD) showing major equipment, valves, and instruments.

- Pressure vessels
- Manifolding—materials of construction
- Instrumentation
- Controls
- Data acquisition and management
- RO skid frame
- Auxiliary equipment

6.1 Cartridge Filters

Cartridge filters are usually used to directly pretreat influent water just prior to the RO membranes. Cartridge filters are designed to prevent resin and media that may have carried over from upstream softeners and filters, from reaching the RO feed pump and damaging the impeller as well as reaching the RO membrane modules and blocking off the feed channels. They are also designed to remove macroparticles that could physically abrade or penetrate the thin membrane layer. Cartridge filters are not intended for bulk removal of suspended solids, turbidity, or SDI (see Chapter 3.8), as is commonly believed. The rating of cartridge filters is usually 5 microns nominal (although absolute is recommended), which is much too large for removal of solids that contribute to turbidity and SDI. Cartridge filters with a rating of 1 to 3 microns absolute are sometime used when colloidal silica or metal silicates are present, but these filters can blind off quickly with bulk solids if the pretreatment prior to the cartridge filter is inadequate and, hence, require frequent replacement. This increases the operating costs for the system due to labor and material expenses.

In addition to the micron rating, cartridge filters are also rated in length using 10-inch equivalents. Cartridge filters can be up to four 10-inch equivalents, or 40 inches in total length. Generally, RO systems use 2.5-inch diameter cartridge filters.

The number of cartridges required is determined by the flow rate of the system. Design protocol calls for a maximum flow rate of about 5 gpm per 10-inch equivalent (2.5-inch diameter) for optimum performance. For example, a 200-gpm flow will require forty, 10-inch equivalents or ten, 40-inch long cartridge filters.

Cartridge filters are generally housed in steel vessels. The largest standard housings hold about twenty-one, 40-inch filters. This housing can handle a flow rate of about 420 gpm. Larger flow rates require multiple housings or a custom housing.

Disposable cartridge filters are recommended over back-washable filters. Back-washable filters suffer from several limitations:

- Risk of breakthrough if the backwashing mechanism fails
- Lower efficiency than disposable filters
- Higher biofouling risk

Disposable cartridge filters should be made of synthetic, non-degradable materials such as nylon or polypropylene. Their construction can be spun-bonded or melt-blown, string-wound, and pleated. These are described below.

- Spun-bonded or melt blown: These cartridge filters are manufactured by thermally either bonding or heat blowing pure polypropylene microfibers so the density is lower on the outside surface and gradually gets denser toward the center of the filter. Spun-bonded or melt-blown filters have high capacity, as particles are trapped throughout the entire cross section of the filter.
- String-wound: These cartridge filters consist of a string of polypropylene (or cotton, nylon, jute, polyester, and so forth) wound around a central core. String-wound cartridge filters rely on Van der Waals forces to capture small particles. These filters suffer from the potential to unload particles at higher pressure drops. Additionally, a slower flow rate is recommended for these filters, about 2–3 gpm per 10-inch equivalent.
- Pleated: These cartridge filters are typically used in higher-purity applications, such as pharmaceutical and microelectronics. The filters can have a multi-layered construction or be single layer. Typical materials of construction include polypropylene, polyethersulfone, and borosilicate glass fiber.

Cartridge filters come with various end cap styles, as shown in Figure 6.3. When replacing used cartridges, they should be replaced with the same end cap style so they will properly mate with the housing.

Upon start-up of new cartridge filters, the initial effluent should be sent to drain. This is necessary to prevent fouling of the RO

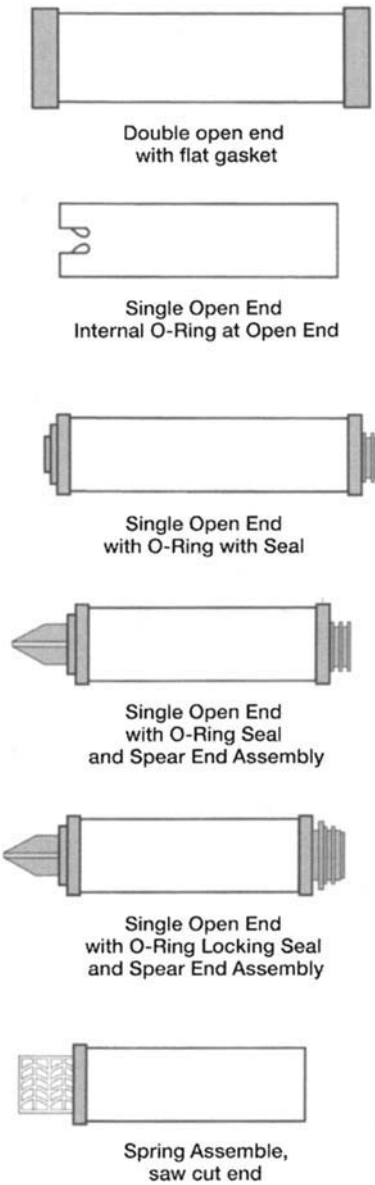


Figure 6.3 Cartridge filter end cap styles. *Courtesy of Siemens Water Technologies, Inc.*

membranes with materials used in the manufacture of the cartridge filter media. For example, lubricants and emulsifiers are used in the manufacture of the strings used in string-wound cartridge filters. These materials can coat the RO membranes and foul them. In the case of polypropylene depth filters, phthalates are used in their manufacture. As little as 50 ppb of phthalate will irreversibly foul an RO membrane.¹

Cartridge filters should be inspected regularly during use to determine the extent of fouling of the filter. Filters should be replaced when the differential pressure across them reaches 5 psi or 2 weeks, whichever comes first. At worst, filters operating with relatively clean feed water (such as RO permeate or well water low in iron) should be replaced at least once per month to minimize biofouling risks.

Inspection of spent filters can yield useful information into the nature of foulants that may be on the RO membranes. Scrapings of the material trapped by the filter can be analyzed for elemental content. The results of this analysis can directly lead to upgrades of the pretreatment system (see Chapter 8).

6.2 Reverse Osmosis Feed Pumps

The most common type of industrial, brackish-water, RO feed pump (sometimes referred to as the “booster” pump) is a centrifugal pump, although some older units still use positive displacement pumps.² Centrifugal pumps are well suited to brackish-water RO applications because these pumps operate favorably at medium flows (typically less than 1,000 gpm) at relatively low pressures (up to 400 psig). Positive displacement pumps have higher hydraulic efficiencies but are plagued with higher maintenance requirements relative to centrifugal pumps.³

Reverse osmosis feed pumps are sized using the required flow rate and operating pressure. Pump curves, as shown in Figure 6.4, are then consulted to determine the number of stages, impeller diameter, and horsepower (hp), required, as well as the efficiency of the pump. For example, a 200 gpm influent flow to an RO that requires 250 psig operating pressure would need a 4-stage pump with a 6.69 in diameter impeller, and an 60 hp motor, after consulting Figure 6.4. The pump motor would operate at 3550 rpm. (Note that this pump will increase the pressure of the influent water by

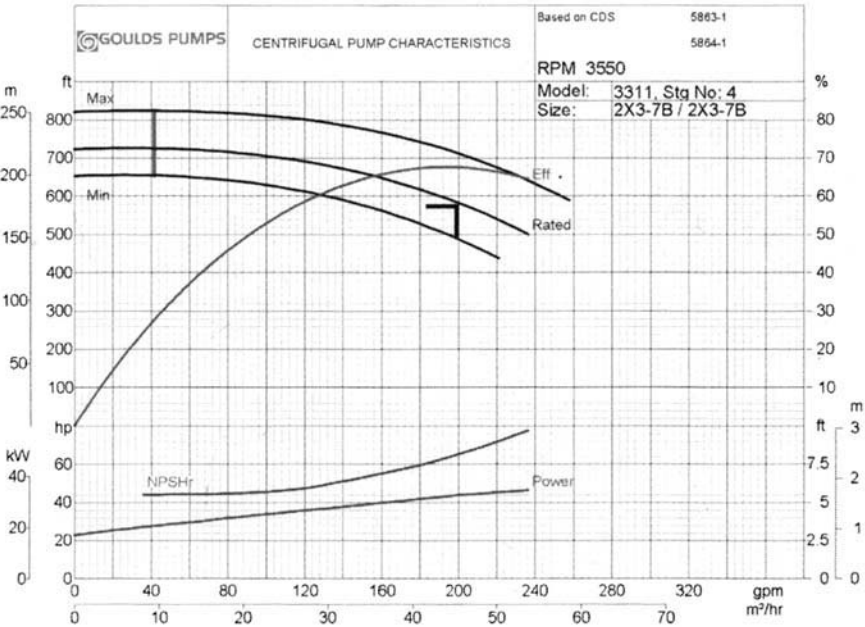


Figure 6.4 Pump curve showing stages, impeller diameter, efficiency, and horsepower. *Courtesy of ITT-Goulds Pumps.*

250 psi over the pump suction pressure. If the suction pressure is 10 psig, the discharge pressure will be 260 psig). The curve also indicates that the net pressure suction head required (NPSHr) to prevent cavitation of the pump is about 3.45 psi (8.0 feet of water). The efficiency of the pump is about 68%, just about the maximum efficiency for this pump. This pump would be quite suitable for use in the specified RO application. However, in the case where the actual pump efficiency was far from the theoretical maximum, another pump would need to be selected that would yield higher pump efficiency. Motor efficiencies run at about 90%. Each pump and motor combination has its own specific pump curve.

Variable frequency drives (VFDs) are sometimes used to adjust the operation of typical (US) standard of 480VAC, 3PH, 60hz operation of the motor. The functionality of a VFD is to convert frequency measured in Hertz (Hz) to motor speed. One Hz equals 1 cycle per second. When voltage is being received (input to the VFD), it is in the sinusoidal waveform. The sine wave is converted to a digital square wave that now controls the revolutions per minute (RPM) of the motor.

The VFD should receive an analog input signal from the permeate flow sensor. This is best practice for utilizing a VFD on an RO system. Some PLC manufacturers also make VFD controllers. This allows for ethernet cable connection from the PLC directly to the VFD. With ethernet cables an analog output is not required to control the VFD. A VFD can also be used with pressure sensors to adjust the RPM. Gaining total control of the RPM's will adjust the speed of the motor to operate at variable pressures (constant throughput) as dictated by the condition of the feed water and the membranes. This is advantageous for RO systems that operate at different water temperatures in the winter and summer. Lower pressures are required at higher water temperatures to produce the same amount of permeate (see Chapter 9.2 for more details on water temperature affects on RO performance). The VFD will reduce the speed of the motor to generate lower discharge pressure to match to lower requirements at higher water temperatures. Energy is, therefore, saved during the warmer summer months when a VFD motor is used.

If the membranes foul or scale, the VFD will automatically adjust the speed of the motor to generate higher pressure to compensate for the fouling or scaling, which lowers flow through the membrane. In this manner, energy and operating cost is conserved at start up and higher energy costs do not come into play until the membranes foul or scale.

An "inverted duty" motor is required for a VFD. Unless the pump in question has this type of motor, it cannot be retrofitted with a VFD.

During operation, adjustments may be made to the VFD, but care should be taken that flow rates and recovery of the RO are not affected. Figures 6.5 and 6.6 show the outside and inside of a VFD control panel, respectively.

The discharge for centrifugal pumps is typically adjusted using a proportional pressure control valve to achieve the required operating pressure (unless a VFD is installed, in which case the pump speed is adjusted to achieved the required discharge pressure). The pressure control valve and concentrate flow control valve (also a proportioning valve) are typically throttled together to achieve the desired productivity and recovery, as shown in Figure 6.7. Manual start-up of an RO unit requires adjusting the concentrate flow control valve and then the pressure control valve, then back to the flow control valve, and so on until the desired performance set-point is reached. The control system will automatically adjust these valves during an



Figure 6.5 Variable frequency drive (VFD) control cabinet. *Courtesy of Nalco-Crossbow Water LLC.*

automated start-up. It is important to always start up the RO with the concentrate flow control valve wide open to prevent damage to the equipment or membranes upon receiving the initial pump discharge pressure blast. (Note that positive displacement pumps are adjusted using a flow control valve on a pump recycle line. The recycle line is necessary because the pump output volume is fairly constant over the entire range of output pressures for this type of pump.³⁾

As membranes age, their performance changes negatively due to the effects of fouling and scaling or degradation. These changes require adjustment in the control valve settings. For example, assuming that membrane flux declines about 15% over three years, the pressure control valve will need to be throttled to increase the discharge pressure to compensate for the loss in flux (hence, the pump must be oversized for initial conditions to allow for this increase in discharge pressure at the same flow rate).² RO feed pumps should be selected based on a 10% pressure premium over



Figure 6.6 Inside of a variable frequency drive (VFD) control cabinet. *Courtesy of Nalco-Crossbow Water LLC.*

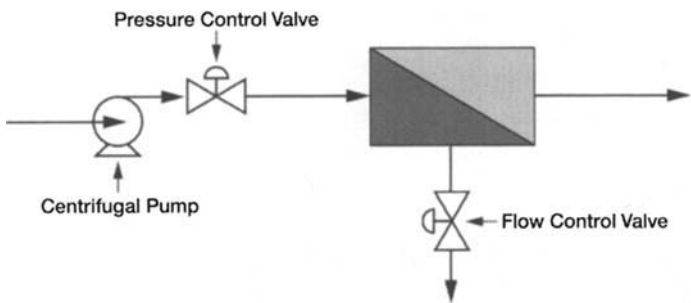


Figure 6.7 Pressure and concentrate flow control valves used to achieve desired productivity and recovery in an RO system with a centrifugal feed pump.

the 3-year membrane life pressure requirements as projected by the RO design programs (see Chapter 10). This insures that enough pressure has been built in to the pump and motor to overcome any irreversible fouling that may occur over the life of the membranes. In such cases, a VFD would save energy and operating cost by only applying the pressure that is actually needed at any given time.

Pumps should be started slowly to prevent water hammer (a surge resulting from a sudden change in liquid velocity). Water hammer can cause cracks in the outer shell of the membrane modules as well as compaction of the membrane itself (compaction results in lower flux through the membrane at constant pressure). Also, water hammer causes the membrane modules to move in the vessel, which can cause wear to the O-rings used on standard interconnectors and lead to leaks of feed water into the permeate (see Chapter 4.3.3). An increase in pressure of no more than 10 psi per second is recommended.³ Some motors may be equipped with a "soft start" that regulates the speed with which they start up. Other considerations to minimize water hammer include:

- Use centrifugal pumps. If positive displacement pumps must be used, these pumps must be fitted with approved pulsation dampening equipment; surge tanks may also be required, particularly for very short and very long pipe runs.
- Air should be vented from the system, either via the flushing cycle or mechanical vents at the uppermost section of the pipe-work in question.
- Valve operation.
 - Flow valves should be open when pumps are activated.
 - During valve change over, the closed valve should complete it's opening cycle before the open valve closes.
 - Flow valves should fail open.
 - Valves should have adequate actuation time. A solenoid valve closing in 40 milliseconds in a stream pressurized to 50 psig will generate a total pressure spike of about 490 psi.⁴

An RO feed pump requires a certain volume and pressure of make-up water to the suction side of the RO feed pump so as not to

cavitate the pump, as discussed above. Low pressure and volume to the suction side of a pump are typically caused by one of the following three problems:

1. Excessive pressure drop through the pretreatment system, including the pre-filter installed on the RO.
2. Deficient pretreatment design. If pretreatment equipment is designed to backwash with service water while the RO unit is on-line, allowances must be made so that enough flow reaches the suction of the RO feed pump while the pretreatment equipment is in backwash.
3. Post installation modifications. Many times, new applications for make-up water are developed within the facility. In some cases, the easiest take-off point location is the make-up pipe to the RO system. However, if the new demand for water is significant, it will starve the RO system.

6.3 Pressure Vessels

A pressure vessel is the pressure housing for the membrane modules and contains the pressurized feed water. Various pressure ratings are available depending on the application:

- Water softening (nanofiltration—see Chapter 16.2): 50 psig up to 150 psig
- Brackish water reverse osmosis: 300 psig up to 600 psig
- Seawater reverse osmosis: 1,000 psig up to 1,500 psig

Pressure vessels are available in non-code or ASME-coded versions.

Pressure vessels are made to specifically to accommodate whatever diameter of membrane module being used, be it a 2.5-inch diameter tap water membrane module up to 18-inch diameter industrial membrane module. The length of the pressure vessel can be as short as one membrane module in length up to seven membrane modules in series (see Figure 6.8).

Most pressure vessels are side-entry and exit for the feed and concentrate, although some older systems employ end-entry and exit vessels. Side-entry pressure vessels are preferred over end-entry vessels because the amount of piping that must be disconnected to

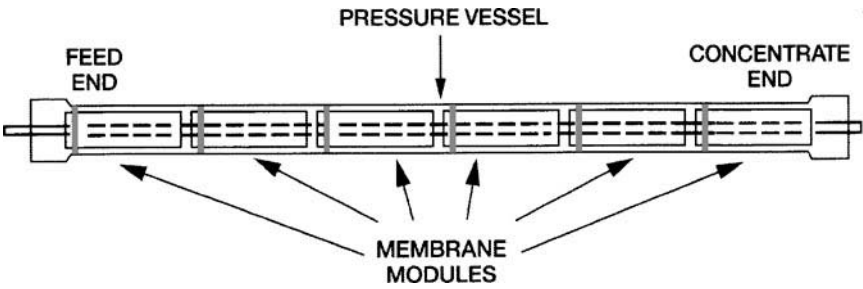


Figure 6.8 Pressure vessel containing six spiral-wound modules housed in series.



Figure 6.9 Pressure vessel showing side entry of feed water. *Courtesy of Nalco-Crossbow Water LLC.*

open the end of a pressure vessel for module replacement is minimized; only the permeate piping must be disconnected. Permeate exits out of the end of the pressure vessel in either configuration. Figure 6.9 shows side-entry pressure vessels.

Proper installation of membrane modules into a pressure vessel is critical. The membrane modules are guided into the pressure vessel in series. Membranes should be loaded into pressure vessel in the direction of flow. That is, the concentrate end of the module (the end without the brine seal) is inserted first into the pressure vessel. The brine seal and O-rings on the module inter-connectors can be

lubricated to ease installation. They should always be lubricated using silicone, glycerin, or water; petroleum products and vegetable-based oils are not recommended as they can damage the module and membranes, thereby voiding the membrane warranty.⁵ Once the first module has been loaded into the pressure vessel, the second module is connected with its concentrate end to feed end of the first module. These two modules are then pushed into the pressure vessel using additional modules that are connected in the same manner (the final module into the pressure vessel may require additional force to push it and the other modules all the way in. Usually, a tap with a mallet on the end of the last module can move the train of modules into place). The first module into the pressure vessel becomes the last one in the series for that stage of the array.

Modules should also be removed from pressure vessel in the same direction as the flow. Hence, the first module into the vessel, which is the last one in the series, is the first module out. Figures 6.10 a, b, and c show the removal sequence for a FilmTec *i*LEC membrane module (see Chapter 4.3.3).

Pressure vessels are usually constructed of fiberglass or stainless steel. Fiberglass is typically used for industrial, non-sanitary applications. Stainless steel vessels are preferred for sanitary applications, where high-temperature (up to 85°C) cleaning performance may be required.

Each pressure vessel is supplied with end caps. Figure 6.11 shows an end cap with an elliptical head with the end adaptor in place. This end cap would be placed at the feed end of the pressure vessel. Figure 6.12 shows an elliptical end cap with the thrust cone (see discussion below) for a side-entry pressure vessel. Figure 6.13 shows a standard end cap with thrust ring for an end-entry pressure vessel. Thrust cones/rings are used on the concentrate end of the pressure vessel. End caps are held in place with a snap ring, shown in Figure 6.14.

Thrust rings/cones and shims are used in conjunction with pressure vessel end caps to minimize longitudinal movement of membrane modules within the pressure vessel. Movement of the membrane modules can cause the O-rings to wear as well as cause telescoping of the membranes and spacers during pressurization. Thrust rings/cones also serve to distribute the axial pressure load to the full end cap.

Correct installation of the thrust ring/cone and shims on the pressure vessel end caps and adaptors is important.

- Thrust ring/cone: the thrust ring/cone is designed to protect the end cap of the last membrane module

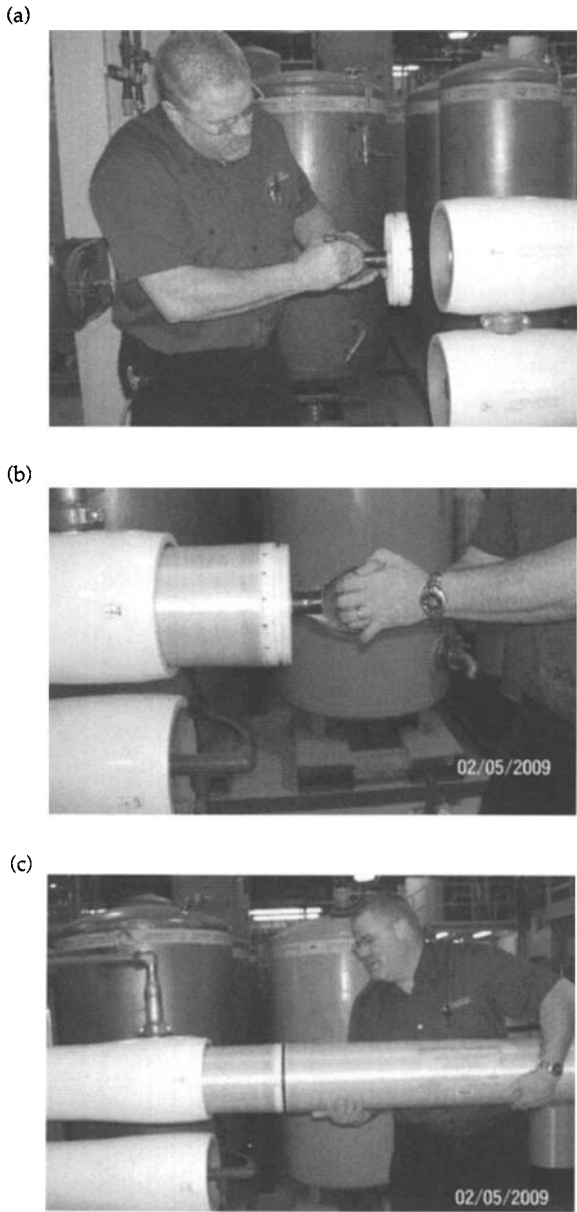


Figure 6.10 Sequence of module removal for Dow Water and Process Solutions-FilmTec *iLEC* membrane modules, a) module removal device, b) pulling module out of pressure vessel, c) disconnecting 2 modules. *Courtesy of Nalco-Crossbow Water LLC.*

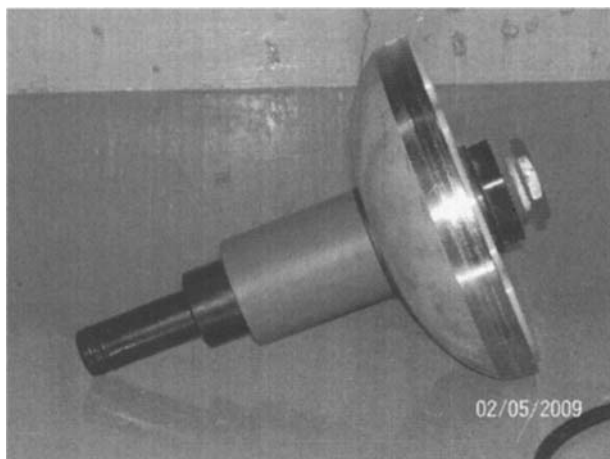


Figure 6.11 Elliptical pressure vessel end cap with end adaptor in place for feed end of pressure vessel. *Courtesy of Nalco-Crossbow Water LLC.*

from being destroyed during pressurized operation. It is installed at the discharge or concentrate end of the pressure vessel. Due to variations in design, the user should consult the specific vendor's specifications on how to correctly position the thrust ring. Figure 6.15 is a photo of a thrust ring as positioned on the elliptical end cap of the pressure vessel (the end adaptor similar to the feed adaptor shown in Figure 6.10 is hidden behind the cone). The thrust cone performs the same service as the thrust ring. However, the cones are easier to install, as the cone separates from the end cap and there is no specific orientation required.

- **Shims:** Shims are used to prevent modules from moving back and forth during pressurization and depressurization. Such movement could wear on the internal O-ring seals. Shims are plastic spacer rings similar to washers. They are typically 0.20-inches thick, and can be purchased from the manufacturer of the pressure vessel or fashioned from polyvinylchloride (PVC) pipe. Shims fashioned from PVC pipe must be cut parallel and free of burrs to work correctly. They are installed between the face of the lead membrane module and the adapter hub (see Figure 6.16) after all the

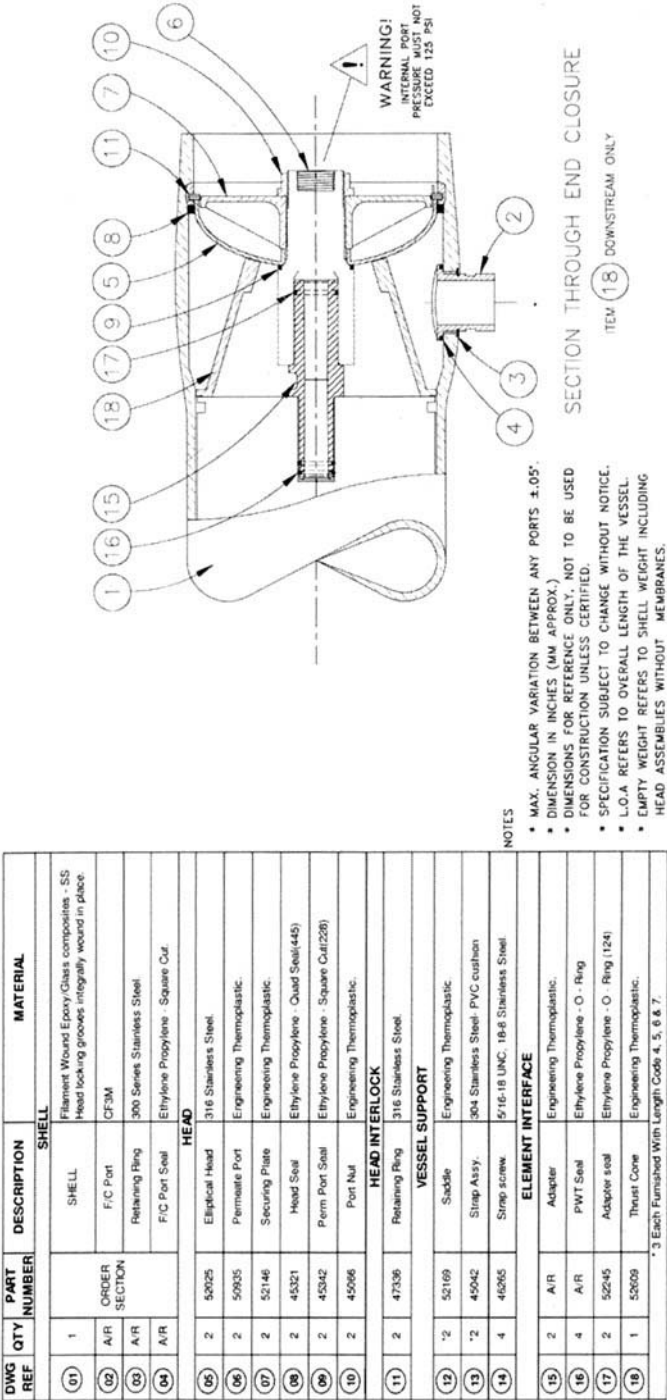


Figure 6.12 Elliptical pressure vessel end cap with thrust cone in place for concentrate end of pressure vessel. Courtesy of Pentair Water.

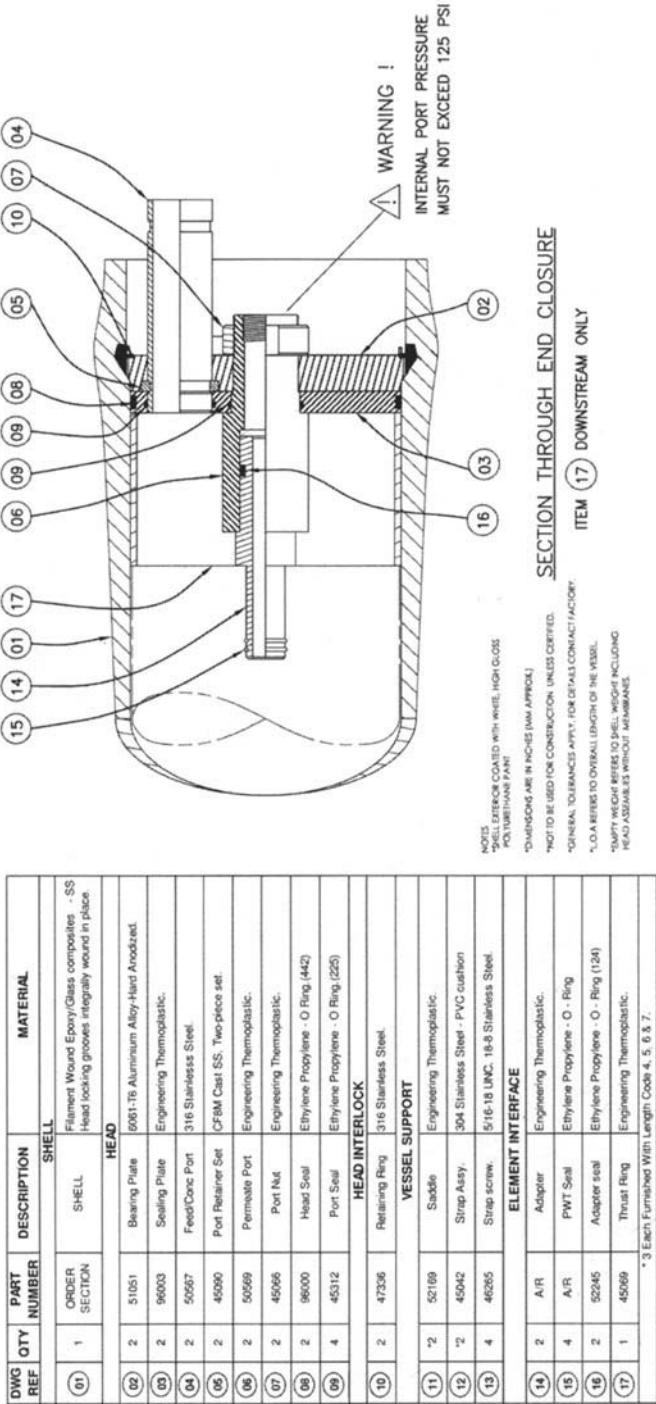


Figure 6.13 Standard pressure vessel end cap with thrust ring for end-entry pressure vessel. Courtesy of Pentair Water.



Figure 6.14 Snap ring that holds pressure vessel end cap in place. *Courtesy of Nalco-Crossbow Water LLC.*

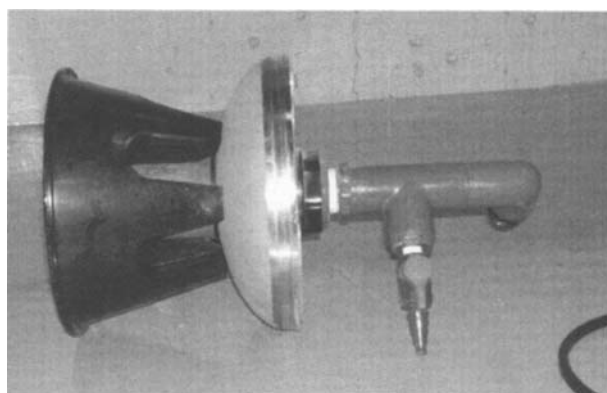


Figure 6.15 Elliptical head with thrust cone in place for concentrate end of pressure vessel. *Courtesy Nalco-Crossbow Water LLC.*

membrane modules have been loaded into the pressure vessel. Prior to installation of the shims, the membrane modules should be pushed completely into the pressure vessel so that the modules seat firmly against the thrust ring.

Prior to installation of the end caps, the head seal should be installed. The head seal is an O-ring to prevent feed/concentrate from leaking out of the end caps. Figure 4.22 shows the location of this O-ring, as do Figures 6.12 and 6.13.

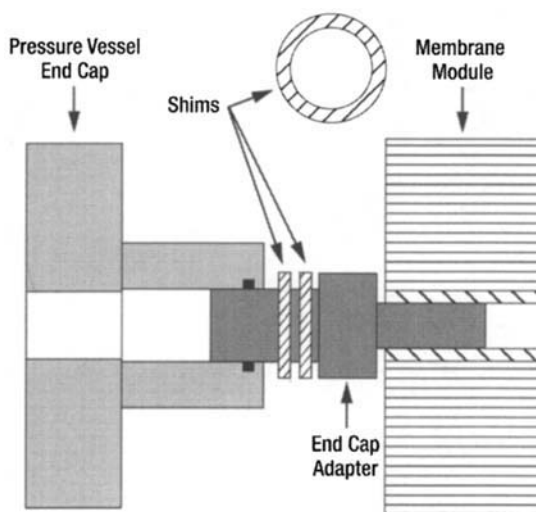


Figure 6.16 Placement of shims between the lead membrane module and the adapter hub. *Courtesy of Dow Water and Process Solutions.*

6.4 Manifolding—Materials of Construction

The low pressure piping on an RO skid is typically schedule 80 PVC. This includes the feed, low-pressure concentrate, and product piping. High pressure piping is typically schedule 10, 316L stainless steel (suitable for waters with concentrate streams below 7,000 ppm TDS). Sanitary applications (such as food, pharmaceutical, or biotechnical processing—See Chapter 4.4.2.6) are generally all stainless to allow for disinfection of the system.

RO permeate distribution piping considerations need to be mindful of the fact that the permeate is highly corrosive. Retrofitting an RO system into a facility with carbon steel permeate piping is difficult, as the piping will corrode. Nonmetallic materials such as plastics and fiberglass are recommended for low-pressure RO product distribution piping.

6.5 Instrumentation

Instrumentation is key to operating and monitoring an RO system. Unfortunately, there is little uniformity among RO equipment vendors in the instrumentation they provide. Table 6.1 lists basic

Table 6.1 Basic recommended RO instrumentation .

Parameter	Raw Feed	Pressurized Feed	Interstage	Product	Reject
Pressure	X	X	X	X	X
Flow	X		X	X	
Conductivity	X		X	X	X
Temperature	X*				
ORP	X*				
pH	X*				

* Typically installed off skid

instrumentation that should be part of each RO system. Most vendors do supply the influent, reject, and permeate instrumentation listed with the exception of the pH, temperature, and chlorine or ORP monitors, which are sometimes available as options. However, many vendors do not include the interstage instrumentation. This is an important omission, as this instrumentation is vital to determining whether problems with an RO system are due to fouling in the first stage of an RO or scaling in the last stage of an RO.

Alarms and shutdowns are necessary to prevent damage to RO and pretreatment equipment and to personnel near the unit. Common shutdown alarms are listed below:

- **Low influent flow**—insufficient flow is available to keep the RO feed pump flooded and operating properly.
- **Low reject flow**—the recovery of the RO system is too high; this shutdown alarm also protects against completely closing the reject flow control valve, which could damage the membranes.
- **Low permeate flow**—not enough flow is available for downstream processes such as polishers.
- **Low permeate quality**—permeate should be diverted as the RO shuts down.
- **Low influent pressure**—this could lead to cavitation of the RO feed pump.

- **High pump discharge pressure**—this is to protect the RO membranes from high-pressure water.
- **High influent temperature**—this is to protect the membrane module materials of construction that can “melt” at high temperatures.
- **High oxidation potential**—this is to prevent membrane degradation by free chlorine or other oxidizers.

6.6 Controls

Most RO skids are equipped with either a microprocessor or programmable logic controller (PLC). Both the microprocessor and the PLC replaced mechanical relay panels, that were very large in size, and had tendencies for difficult troubleshooting. From the early time of RO manufacturing, control panels in most cases were large enough for the average-sized human to stand in. Today's technology allows for controls to be mounted directly to the RO units, and save a great deal of space. The PLC and microprocessor offer digital relay technology that are connected within a base moduals, other wise known as bricks (or chipsets). This is opposed to the electromechanical relay.

Microprocessors are usually found on smaller or lower-priced RO systems, while PLC controls are used for larger, more complicated systems that require greater control over process conditions. Major suppliers of PLC units for RO systems include Allen-Bradley, and Siemens.

Microprocessors boards are used on smaller RO units that require minimum input/output (I/O) function. In general, standard manufactured microprocessor boards, inexpensive are used. Troubleshooting a microprocessor board can be difficult without proper documentation and experience. In many cases, it would be quicker and more cost effective to replace a mal-functioning control board. Damage is usually caused by human errors on field wiring.

The PLC provides more expanded options for control and changes to control. They are provided with many different (I/O) cards such as, digital, analog, device net, modbus, and Internet Protocol (IP). Along with a human machine interface (HMI), the combination makes a solid control system for water treatment. High end PLC RO systems offer pretreatment control, along with multiple external valve option, post-treatment DI, and external pumps.

Systems that are fully automated also include several proportional plus integral plus derivative (PID) controllers. The PID controllers control individual set-point functions and can monitor alarm conditions. Independent PID controllers can control their specific function without a PLC. Should an independent PID controller fail, only the specific function it controlled cannot be adjusted.

Larger, commercial installations will also be equipped with a Supervisory Control And Data Acquisition system (SCADA). The SCADA system is connected via a data highway with the PLC(s) used in the system.

The basic function of a control system is to keep the RO system functioning and on-line. At start-up, the control system will automatically adjust the pressure and flow control valves to achieve the desired productivity and recovery of the system (provided the system is in automated start-up mode). It is highly recommended that an trained RO specialist be on-site for start-up. This will ensure that manufactures warranties are not compromised by human errors. The control system will also automatically turn RO skids on and off according to the demand for product water. This is typically accomplished via level control in the permeate tank. On/off (optional) divert valves can be automatically controlled to divert unsuitable feed water from the RO membranes. Examples of when divert valves spring into action include high temperature, high pH and free chlorine present in the feed water. There may also be product divert valves to dispose of high-conductivity permeate. Waste flush valves can cycle water in between long down time to prevent bacteria growth (see Chapter 13.1).

Control systems also include features to protect the membranes and the feed pump. Pressure switches are used on the pump suction for low inlet pressure and on the discharge for high pressure. A pressure-relief valve is installed on the permeate line to prevent backpressure from damaging the membranes. Back pressure shouldn't be considered an option for the membranes. Some systems may have a high membrane differential pressure switch to prevent crushing of the membranes during conditions of excessive membrane fouling or scaling.

Administrative functions in the PLC are typically set at the factory or during installation by factory-trained personnel. These functions should not be adjusted except by those trained to do so. Changing the functions without understanding how all components of the RO system are interdependent can be disastrous for operation of the RO.

Reverse osmosis systems operate best when the flows and, therefore, recovery of the system are held constant. Both were designed with attention to other variables such as Beta, minimum velocity through a pressure vessel, and maximum velocity through a pressure vessel (see Chapters 3.5, 9.4, and 9.5). These variables and others are important to minimize fouling and scaling of the RO membranes; adjustment without regard to all of the other engineering design considerations will lead to accelerated fouling, scaling, cleaning frequencies, and membrane replacement.

When operating conditions change, for example, when the feed water temperature decreases or membrane fouling occurs, performance such as the permeate flux also changes. Adjusting the feed pressure compensates for such changes in performance. This adjustment can be manually initiated, but generally occurs automatically through the PLC if such a unit has been purchased. Careful observation of the system is required to ensure that the maximum allowable feed pressure is not surpassed or that fouling does not become excessive.

There is a tendency to want to increase throughput shortly after start up or after a successful membrane cleaning, when membranes are performing their best. However, if changes are made without regard to consideration of the other variables in the system that depend on flow and recovery, that will hasten fouling and scaling as a result.

6.7 Data Acquisition and Management

An operator interface is used to record data gathered by the PLC.¹ The operator interface is usually another computer (sometime called the human-machine interface or HMI). The HMI uses process displays with real-time sensor readings so that the operator can quickly assess the status of the system (see Figure 6.17). The operator uses the control panel to adjust alarm settings and to turn on and off process equipment. Once running, however, the PLC controls and runs the system automatically, without further input from the operator. Common HMI status indicators are listed below:

- All shutdown alarms
- Total run time
- RO operating mode



Figure 6.17 Human-Machine Interface (HMI) showing status of the reverse osmosis system, including pretreatment.

- Recovery
- Influent flow
- Reject flow
- Permeate flow
- Pump status
- Valve status
 - Influent
 - Reject
 - Permeate to tank
 - Permeate to drain
 - Permeate flush

Data management typically means normalization. Normalization software is available from several sources to help interpret RO operating data (see Chapters 11.3 and 12 for more detail about “normalized” data). Membrane manufacturers have normalization software that requires manual input of operating data, but that run the calculations. Some chemical and equipment vendors have software/hardware packages that collect data and perform all the normalization calculations automatically. In general, RO control packages and PLCs do not come with normalization software.

6.8 Reverse Osmosis Skid

Reverse osmosis skids are typically contained within a frame of galvanized or urethane-coated steel. Skids should be designed for easy access for monitoring and maintenance. Access to controls, instruments, valves, the pump and motor, and membranes is essential. Access to the permeate from each pressure vessel is often overlooked. Without such access, profiling and probing used to troubleshoot poor performance is not possible (see Chapter 14.7).

When locating space for an RO skid, attention should be given to space needed for access to controls and the membranes. Multiple skids are usually put in face to face. Four feet should be allowed between the face on a skid and neighboring equipment (one to two feet is acceptable for the back side of the skid). Each end of the skid should have at least four feet and preferably six feet of free space so that membranes can be installed and removed from the pressure vessels.

There are advantages to hard piping most of the RO system, but some piping needs to remain open so that flows can be observed and measured. In particular, waste flows to drain, such as filter backwash waste, RO reject, and RO permeate divert streams, should be accessible before they enter the drain so that they can be easily observed and sampled if needed; waste-flow piping should end 6 to 10 inches above the drain.

6.9 Auxiliary Equipment

Some RO skids come equipped with “on board” integrated cleaning equipment. The main process pump and cartridge filter are used for the cleaning system. The skid also includes valves and hoses that are used for the cleaning. A free-standing cleaning tank is placed next to the RO skid. See Chapter 14.2 for more details about cleaning membranes.

If an RO skid does not come with an integrated cleaning system, a free-standing cleaning skid is required (alternatively, the membranes must be shipped out for off-site cleaning—see Chapter 13.2). The cleaning skid should include a cartridge filter housing, low-pressure recirculating pump, and tank. The tank may or may not be equipped with a mixer and heating element. If a mixer and heating

element are not included, the cleaning solution must be recirculated within the cleaning skid until the cleaning chemicals are well mixed and the recirculation pump imparts enough heat to the cleaning solution that an effective cleaning can be conducted.

Product blending “kits” are available by some equipment vendors for factory installation. These kits allow for bypass of a prescribed amount of influent water into the permeate line. The kit includes check and butterfly valves, a rotameter, and all fittings.

Automated SDI kits are available that can be installed on the RO influent line (see Chapter 3.8 for more information about SDI). These kits provide the hardware and software to run up to 5 consecutive SDI tests and perform the SDI calculations without operator attention. Should an automated kit not be available, manual SDI equipment should be installed. Equipment required for a manual SDI system is shown in Figure 3.6.

6.10 Other Design Considerations

6.10.1 Access to Profile and Probe RO Membranes

Profiling and probing are two techniques use to sample the performance of individual membranes *in situ* (see Chapter 14.7). Access for profiling and probing is important to assist with troubleshooting an RO system. Profiling requires that permeate sample port be installed on the effluent from each pressure vessel. Probing requires that the sample port be such that a section of flexible tubing can be snaked down the through the port into the permeate tubes of the membranes while installed in the pressure vessel. Many commercially-available standard skids do not include proper valves to allow for either profiling or probing.

6.10.2 Interstage Performance Monitoring Instrumentation

Data normalization over individual stages is important to allow for determination of the type of fouling or scaling that is occurring and where in the system is it occurring. Interstage instrumentation that is required includes flow indicators, pressure sensors, and conductivity meters.

6.10.3 Stage-by-Stage Membrane Cleaning

The ability to clean each stage individually in an RO system is very important. This prevents scale from the last stages to be mixed with the other stages, and foulants from the first stage being mixed with other stages. It also allows for proper flow rates through the pressure vessels to maximize cleaning efficacy. Valves should be installed to allow for cleaning of each stage individually. Additionally, if multiple skids are cleaned with the same cleaning skid, provisions should be made to allow return piping to drain completely between cleaning of each RO skid. This will eliminate cross contamination from one skid to the next.

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PRETREATMENT

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Water Quality Guidelines

The performance and successful operation of an RO system depends directly on the quality of water feeding the RO. The nature of feed water constituents can influence membrane performance by causing scaling, fouling, or degradation of the membrane. Table 7.1 lists water quality guidelines against which RO influent (and concentrate) should be contrasted, to determine whether membrane fouling, scaling, or degradation is possible. This chapter details various feed water constituents that affect the performance of RO membranes.

7.1 Suspended Solids

Suspended solids are typically measured using turbidity. Turbidity measures the light-scattering ability of particles in water. The water quality guidelines call for an influent turbidity of less than 1 Nephelometric Turbidity Units (NTU), which also happens to be a warranty requirement of membrane manufacturers. Exceed 1 NTU and the membrane warranty is voided. The lower the turbidity, the less likely the membranes are to foul with suspended solids. RO best practices call for feed water turbidity less than 0.5 NTU.

Another measure of suspended solids is particle size distribution. However, there are no recommendations on particle size distribution in RO feed water that have been established.

Silt density index measures suspended solids, particularly colloids, such as alumina- or iron silicates, clay, iron corrosion products, and microbes, that have a great potential for fouling RO membranes (see Chapter 3.8 for more details about SDI). The SDI should be as low as possible to minimize fouling of the membranes, but must be less than 5 to meet warranty requirements set by the membrane manufacturers (best practices call for SDI in RO feed water to be less than 3). Note that there is no direct correlation of turbidity to SDI, other than high turbidity usually means high SDI (the converse is not always true).

Table 7.1 Generally-accepted water quality guidelines for RO influent and concentrate waters.

Species	Units	Guideline Value/Range
Colloids	SDI (unit-less)	< 5*
Suspended Solids	NTU	< 1
Calcium Carbonate	LSI	< 0**
Metals: iron, manganese, aluminum	ppm	< 0.05
Barium, Strontium	ppm	< 0.05
Hydrogen Sulfide	ppm	< 0.1
Microbes	CFU/ml	< 1,000†
Silica (soluble)	ppm	140 – 200**
Organics (TOC)	ppm	< 3
Color	APHA	< 3
Chemical Oxygen Demand (COD)	ppm	< 10
pH—CA Membranes	pH units	4 – 6
pH—PA Membranes	pH units	2 – 12***
Chlorine, free—CA Membranes	ppm	< 1
Chlorine, free—PA Membranes	ppm	< 0.02
Temperature—CA Membranes	°C	< 30
Temperature—PA Membranes	°C	< 45

* Silt density index (see Chapter 3.8)

** Can be up to 2.0 – 2.5 depending of the type of antiscalant used

†In RO reject stream

**In RO reject stream, varies as functions of pH and temperature (see Figures 7.1 and 7.2)

***General guideline—check with membrane manufacturer for limits for specific membranes

Membranes fouled with suspended solids will exhibit lower productivity and an increase in pressure drop. Sometimes there is also a decrease in salt rejection.

Suspended solids can be removed or reduced in RO feed water using coagulation, clarification, and filtration (see chapter 8.1.1).

7.2 Microbes

Microbial fouling of RO membranes is a significant issue. Bacterial colonies will grow virtually anywhere in the membrane module where the conditions are favorable. Concentration polarization provides an environment next to the membrane surface that is enriched in nutrients for microbes. Satellite colonies can break off and begin to grow elsewhere within the membrane module, increasing the surface area of membrane that is covered with microbes and their associated biofilm. For a more complete discussion on membrane biofilms, see reference 1. Microbial fouling will lower membrane productivity, increase operating pressure, and increase pressure drop.

The potential for biological fouling of a membrane can be determined by considering the assimilable organic carbon (AOC). This test is a bioassay that measures the growth potential of microorganisms in a sample. The test procedure is described in part 9217 of the Standard Methods.² A value of 10 micrograms per liter is a proposed standard to minimize biological fouling, but in some cases, fouling may still occur even at this low value.³ Dow Water Solutions recommends an AOC value of less than 5.⁴

The degree of membrane fouling with microbes that has already occurred is determined by checking the number of colonies that slough off the membrane into the RO reject stream. This is typically determined using one of two methods:

- **Culture:** This technique is easy to perform and does not require expensive equipment. It is used to determine the number of colony-forming units (CFU) in a water sample using part 9000 of the Standard Methods³. The number of CFU in a sample is an expression of the number of culturable microorganisms present. Note that while this technique is relatively inexpensive, the counted colonies may represent only about 1 – 10% of the total bacterial count (TBC). Never-the-less, this technique can be useful

in tracking microbial fouling. Concentrations of 1,000 CFU per milliliter or greater in an RO concentrate stream are considered a fouling problem that can significantly and negatively effect performance of the RO system.

- **Total Bacteria Count:** The TBC is determine by directly counting the actual number of microorganisms collected on a filter after it is used to filter a sample of the water in question.⁵ The sample is stained with acridine orange and viewed with an epi-illuminated fluorescent microscope. This technique is more accurate and quicker than the culture technique, but is not as practical for field work.

Microbial fouling is best dealt with before biofilm becomes mature. Biofilm protects the microorganisms from the action of shear forces and biocidal chemicals used to attack them. Microbes can be destroyed using chlorine, ozone, ultraviolet radiation, or some non-oxidizing biocides (see Chapters 8.2.1, 8.2.2, 8.1.8, and 8.2.5, respectively). An effective method to control bacteria and biofilm growth usually involves a combination of these measures. Specifically, chlorination or ozonation of the pretreatment system, followed by dechlorination to protect the membranes, or UV distruction followed by periodic sanitation with a non-oxidizing biocide used directly on the membranes.

7.3 Organics

Organics adsorb to the membrane surface resulting in flux loss that can be permanent in some cases.⁴ Adsorption is favored at pH less than 9 and where the organic compounds are positively charged. Particularly troublesome are emulsified organics, which can form an organic film on the membrane surface. Organic fouling exacerbates microbial fouling, as many organics are nutrients for microbes. It is recommended that the organic concentration, as measured by total organic carbon (TOC) be less than 3 ppm to minimize fouling potential. Organic fouling of the membrane will decrease productivity of the membrane.

The concentration of oils (both hydrocarbon and silicone-based) and greases should be less than 0.1 ppm in RO feed water. These materials will readily adsorb onto polyamide membranes and result in a decrease in membrane throughput. However, they can

be removed from the membrane using alkaline cleaners if the flux has not declined by more than 15% from start-up.⁴

Organics can be reduced in RO feed water using coagulation/clarification, ultraviolet radiation, or activated carbon filtration (see Chapters 8.1.1, 8.1.8, and 8.1.4, respectively). Lower molecular weight organics such as urea, iso-propyl alcohol, and acetone are not easily removed using these techniques. However, oxidation of these organics using persulfate activated by UV light has been shown to be successful.⁶ Oils and greases can be removed from RO feed water using coagulation/clarification, carbon filtration, with special carbon or, in the case of wastewater reuse, dissolved or induced air flotation.

7.4 Color

Color also adsorbs onto the surface of the RO membrane. Color is typically made up of naturally occurring humic substances that form when organic substances such as leaves decay. Humic substances are themselves composed of three different types of organic compounds. Humic acid is that color which precipitates during acidification; these organics are dark brown to black in color. Fulvic acid does not precipitate during acidification; these substances are yellow to yellow-brown in color. Finally, humin is not soluble at any pH and is black in color.

Color can be true or apparent. Apparent color is essentially total color, composed of dissolved and suspended organics and other suspended solids such as iron oxidizes. True color is measured by filtering out the suspended matter so that the only color present is due to dissolved organics. Color is measured using APHA (American Public Health Association) dimensionless units.

Adsorption of color onto an RO membrane is favored when the compounds are hydrophobic or positively charged. As with other organics, a high pH (>9) helps to minimize fouling with color, but causes other concerns, including calcium carbonate scaling.

True color should be less than 3 APHA to minimize fouling due to color adsorption. Adsorption of color onto the membrane will decrease productivity of the membrane.

Color can be reduced in RO feed water using coagulation/clarification with hydroxide flocculants, ultrafiltration and nanofiltration, adsorption of activated carbon, and ultraviolet radiation (see Chapters 8.1.1, 8.1.9, 8.1.4, and 8.1.8, respectively).

Note that exposure to chlorine can result in color forming trihalo-methanes (THMs), which are known to possess carcinogenic properties (see Chapter 8.1.9 for more information about THMs and chlorination). This is a particular concern to potable or municipal RO systems.

7.5 Metals

RO membranes will readily foul with precipitated metals, including iron, manganese, and aluminum. Soluble iron and manganese (and cobalt present in some bisulfite solutions used for dechlorination) are also a problem for RO membranes. These metals will catalyze the oxidation of the RO membrane resulting in degradation of the membrane. By dropping the pH and reducing the oxygen concentration, higher concentrations of soluble iron can be tolerated. Metal fouling will increase pressure drop and decrease productivity. Oxidation of the membrane with soluble metals will result in lower salt rejection and higher productivity.

Iron and manganese are naturally occurring in well water, although they can be found in surface waters in lower concentrations (the exceptions include mine drainage water). Typically, iron and manganese will be soluble while in the well, but upon exposure to oxygen in air, they precipitate, forming oxides. These oxides collect on the membrane surface, fouling the membrane.

Iron and manganese can be removed from RO feed water using sodium softening or iron filters (see Chapters 8.1.6 and 8.1.5, respectively). In some cases, it may be desirable to operate with soluble iron and/or manganese through the RO system, if the entire system can be kept air tight to prevent oxidation of the metals into suspended solids.

Carry-over and overfeed of alum (aluminum sulfate) coagulants, post-precipitation of alum coagulants due to poor pH control, and naturally-occurring aluminum silicates are responsible for aluminum fouling of RO membranes. Alum feed is usually employed on surface waters where clarification is the first pretreatment step. The overfeed of alum occurs when the raw water turbidity increases. Many operators will continue to add alum past the point called for by stoichiometry. Carry over occurs when the clarifier is not operated properly.

Aluminum can react with silica in water to form aluminum silicates. This reaction can occur at silica concentrations as low as 10 ppm.⁷

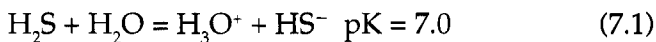
Alum's minimum solubility is at pH 6.5. If the RO is run at pH 7–9, this should keep the alum in solution through the RO. Alum

is addressed either by operating at appropriate pH (7–9), or by replacing the alum with another coagulant.

7.6 Hydrogen Sulfide

Hydrogen sulfide is typically found in well water that is devoid of oxygen. This compound easily oxidizes and releases elemental sulfur, which is very sticky and results in irreversible fouling of RO membranes. Metal sulfides can also form, which can precipitate. Deposits can be sooty-black or a pasty-gray.⁸ Fouling with elemental sulfur or metallic sulfides will cause a decrease in flux and an increase in salt passage.

The following equations show the dependence of the amount of hydrogen sulfide in solution on the pH of the system:



Treatment recommendations are polar opposite for hydrogen sulfide. Some professionals recommend that no treatment to remove the compound be used and all efforts should be focused on keeping the RO feed water free of oxygen or other oxidizers. It may be possible to operate with hydrogen sulfide present *IF* the feed water can be kept free of oxygen or any other oxidizing agents to avoid oxidizing the compound and forming elemental sulfur or metal sulfides. Any back flow of water into the well will create a vacuum on the system which can lead to system voids where air can enter the system and oxidize the hydrogen sulfide. Submersible pumps with check valves located on the discharge of the pump or a procedure to waste the initial flow from the well are techniques that can be used to minimize the potential for fouling with sulfur.

Others recommend treatment to remove the compound. Hydrogen sulfide can be removed or reduced from RO feed water using some iron filters (such as manganese greensand and Filox, see Chapter 8.1.5) or a complicated combination of oxidation, coagulation and filtration, sulfite addition, and rechlorination.⁹

Because hydrogen sulfide is a gas, it is not rejected by an RO membrane. For systems that do not remove the compound prior to the RO membranes, hydrogen sulfide will be present in the permeate. Exposure to air on the permeate side of the membrane, such

as during shut down of the system, can result in an ivory to yellow precipitate on the permeate side of the membrane. This will result in a loss of flux (increase in operating pressure) over time. It is not possible to remove the precipitate manually. However, due to the corrosive nature of the RO permeate, standard operation of the RO system will eventually remove the precipitate, if additional fouling with sulfur is prevented. Keeping the feed water pH below 6, as shown in Equation 7.1, can help to keep elemental sulfur and metal sulfide from forming on either side of the membrane.

7.7 Silica

Silica, as insoluble silicates and as soluble or "reactive" silica, can cause problems for an RO system. Insoluble silicates form when silica precipitates. When iron and aluminum are present, silicates of

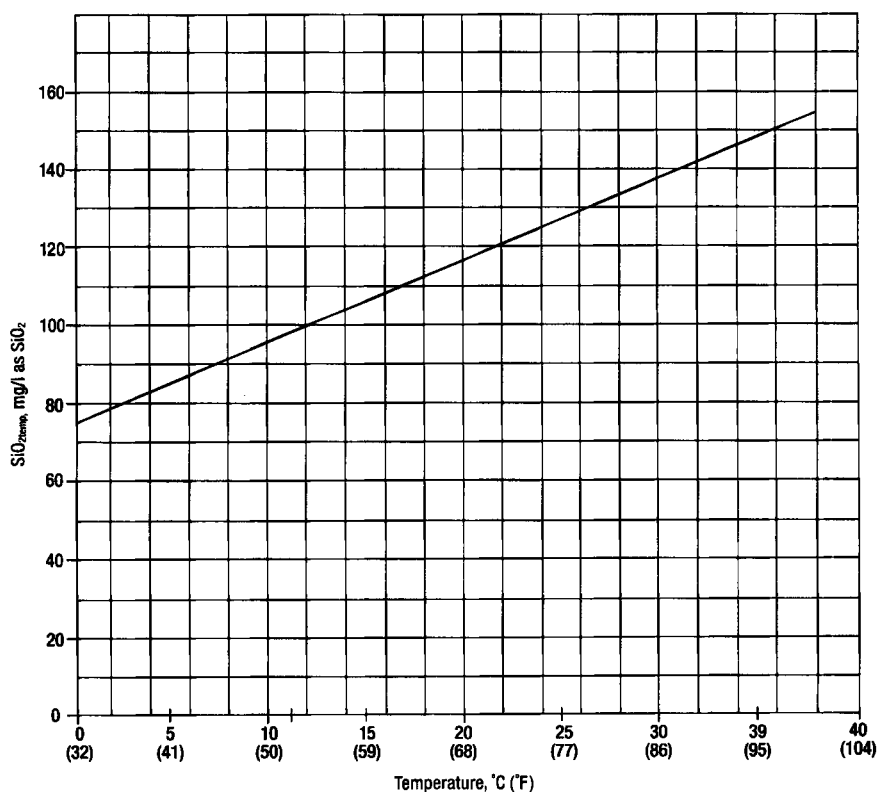


Figure 7.1 Silica solubility as a function of temperature.

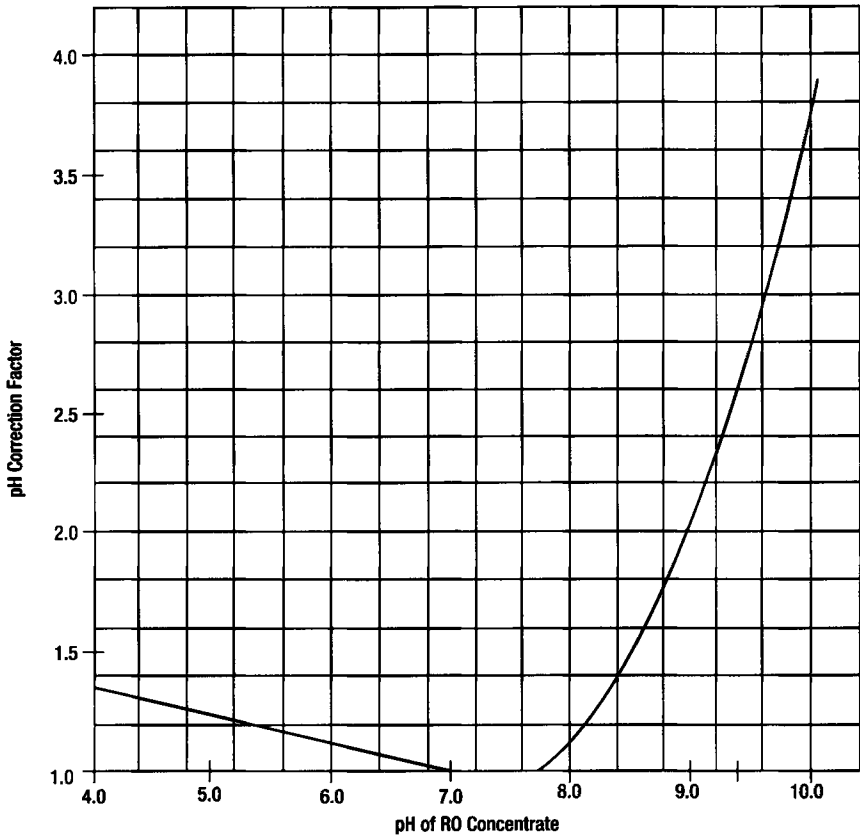


Figure 7.2 Silica solubility as a function of pH. To determine silica solubility at a given pH, multiply the solubility as a function of temperature by the pH correction factor of the given pH of the concentrate solution.

these metals can form quickly and at silica concentration less than saturation.⁴ Saturation of soluble silica is a function of temperature and pH. Silica is more soluble at higher temperature and at pH below 7.0 and above 7.8 (See Figures 7.1 and 7.2). High Efficiency Reverse Osmosis (HERO™ is a trademark of Debasish Mukhopadhyay) is based on these facts; high recovery of high-concentration silica solutions is possible at high pH (see Chapter 16.4).

Soluble silica often limits the recovery of an RO system because of the potential for scaling and the difficulty in removing silica scale from membranes. Silica antiscalants that can handle up to about 200 ppm silica (depending on the conditions and antiscalant manufacturer) are available.

Silica fouling causes high pressure drop and low productivity. Silica scaling causes low rejection of silica and perhaps other ions as well.

Colloidal silica can be removed or reduced in RO feed water using coagulation/clarification (see Chapter 8.1.1). Soluble silica is typically dealt with using high pH, elevated temperature, antiscalants (see Chapter 8.2.3), or the HERO™ process. Sometimes, the recovery of the RO system is adjusted to minimize the potential for forming silica scale.

7.8 Calcium Carbonate

Calcium carbonate scaling is perhaps the most common type of problem, with the possible exception of microbial fouling, that RO membranes experience. Fortunately, it is fairly easy to detect and handle. Basically, if the ion product (IP) of calcium carbonate in the RO reject is greater than the solubility constant (K_{sp}) under reject conditions, then calcium carbonate scale will form. If $IP < K_{sp}$, scaling is unlikely. The ion product *at any degree of saturation* is defined as:

$$IP = [\text{cation}]^a[\text{anion}]^b \quad (7.3)$$

where:

IP = ion product

[cation] = cation concentration

[anion] = anion concentration

superscripts:

a = quantity of cation within the salt

b = quantity of anion within the salt

The solubility product *at saturation* is defined as:

$$K_{sp} = [\text{cation}]^a[\text{anion}]^b \quad (7.4)$$

where:

K_{sp} = solubility product

Langelier Saturation Index (LSI) is used to determine the scaling potential of calcium carbonate. (Note that LSI is used up to about 4,000 ppm TDS; higher concentrations rely on the Stiff-Davis Saturation Index.) The LSI is calculated using the following formulas

(note that the concentrate concentration is used to calculate the LSI, as this is where the concentrations of solutes is the greatest):

$$\text{LSI} = \text{pH} - \text{pH}_a \quad (7.5)$$

where:

$$\text{pH}_a = (9.30 + A + B) - (C + D) \quad (7.6)$$

where:

$$A = (\log_{10} [\text{TDS}] - 1) / 10$$

$$B = -13.12 \times \log_{10} (^{\circ}\text{C} + 273) + 34.55$$

$$C = \log_{10} [\text{Ca}^{2+}] - 0.4, \text{ where } [\text{Ca}^{2+}] \text{ is in ppm as } \text{CaCO}_3$$

$$D = \log_{10} [\text{alkalinity}], \text{ where } [\text{alkalinity}] \text{ is in ppm as } \text{CaCO}_3$$

A positive LSI means that scaling is favored; a negative LSI means that corrosion is favored. It is desirable to keep the LSI near zero (or below) in the RO concentrate to minimize calcium carbonate scaling. This is usually accomplished by feeding acid to lower the pH or feeding an antiscalant (see Chapter 8.2.3). Care must be given if sulfuric acid is used to adjust the pH, as this may exacerbate sulfate-based scales, such as calcium sulfate, barium sulfate, and strontium sulfate.

Alternatively, antiscalants can be used to control calcium carbonate scale at LSI values as high as 2.0–2.5, depending on the specific antiscalants. Calcium also forms scales with fluoride, sulfate, and phosphate. The LSI will not help predict these scales; analysis of water quality, using the ion product and solubility constants, is required to determine the potential for scaling with calcium fluoride or calcium phosphate. Antiscalants currently available can address calcium fluoride and calcium sulfate scale; they do not address calcium phosphate scale (although newer antiscalants will be available in the near future to address this scale).

Scaled membranes exhibit lower productivity and lower salt rejection. This lower salt rejection is a function of the concentration polarization phenomenon (see Chapter 3.4). When membranes are scaled, the surface of the membrane has a higher concentration of solutes than in the bulk solution. Since the membrane rejects when the membrane “sees,” the passage of salts will be higher, even though the absolute or true rejection stays constant.

Calcium is removed from RO feed water using sodium softening, or reduced using lime softening (see Chapters 8.1.6 and 8.3, respectively). Dropping the LSI using acid is used to address calcium scaling without removing or reducing the concentration of calcium. Antiscalants are also used to address the issue without reducing the amount of calcium present (see Chapter 8.2.3).

7.9 Trace Metals—Barium and Strontium

Barium and strontium form sulfate scales that are not readily soluble. In fact, barium is the least soluble of all the alkaline-earth sulfates. It can act as a catalyst for strontium and calcium sulfates scale.⁴ Analyses of the ion product with the solubility constants for barium and strontium sulfates is necessary to determine the potential for scaling with these species. If the ion product (IP) for barium sulfate exceeds the solubility constant, scale will form. Note that in the case of strontium sulfate, if $IP > 0.8K_{sp}$, scaling is likely. However, the induction period (the time it takes for scale to form) is longer for these sulfate-based scales than it is for calcium carbonate scale.

Barium and strontium can be reduced in RO feed water using sodium softening (see Chapter 8.1.6). Antiscalant can be used to control or inhibit scaling without reducing the concentration of either species (see Chapter 8.2.3).

7.10 Chlorine

Polyamide, composite membranes are very sensitive to free chlorine (recall from Chapter 4.2.1 that cellulose acetate membranes can tolerate up to 1 ppm free chlorine continuously). Degradation of the polyamide composite membrane occurs almost immediately upon exposure and can result in significant reduction in rejection after 200 and 1,000-ppm hours of exposure to free chlorine (in other words after 200–1,000 hours exposure to 1 ppm free chlorine). The rate of degradation depends on two important factors: 1) degradation is more rapid at high pH than at neutral or low pH, and 2) the presence of transition metals such as iron, will catalyze the oxidation of the membrane.

The mechanism of degradation is the loss of polymer cross-linking. This results in the membrane polymer dissolving, similar to a nylon stocking when exposed to chlorine bleach. Damage is irreversible and will continue as long as the membrane is exposed to the oxidizer.

Chloramines also pose a risk to polyamide, composite membranes (see Chapter 8.2.1.1). Chloramines are virtually always in equilibrium with free chlorine. Although the tolerance of the FilmTec FT30 membrane to chloramines is 300,000 ppm-hrs, FilmTec still recommends that influent water with chloramines be dechlorinated prior

to the membrane.⁴ In most cases, ammonia is added to chlorine to generate chloramines. This leaves open the possibility that there is still some free chlorine available. The most successful chloramine applications seem to be found in wastewater systems with a resident concentration of ammonia, to which chlorine is added to make the chloramines.

Another note of caution with chloramines is the need for good pH control. If the pH gets up to 9, dissolved ammonia gas is present as $\text{NH}_3(\text{g})$, which swells at least some polyamide composite membranes. This swelling can be enough to drop the salt rejection from 98% down to about 85%.¹⁰ Dropping the pH back to about 7 converts the ammonia gas to ammonium ion, which does not swell the membrane and rejection returns to nominal.

The use of chlorine dioxide is not recommended for use with polyamide, composite membranes.⁴ This is because free chlorine is always present with chlorine dioxide that is generated on site from chlorine and sodium chlorate (see Chapter 8.2.1.1).

Initially, polyamide composite membranes that have been degraded due to chlorine attack will exhibit a loss in flux.⁴ This drop in flux is followed by an increase in flux and salt passage.

Chlorine can be removed from RO feed water using sodium bisulfite or carbon filtration (see Chapters 8.2.4 and 8.1.4, respectively). As discussed in Chapter 8.1.4, carbon in carbon filters can aide the growth of microbes so carbon filtration is typically not recommended for dechlorination of RO feed water unless the concentrations of organics is high enough to warrant its use, or if the dosage of sodium bisulfite is too low for accurate control.

7.11 Calcium

Besides calcium carbonate, there are three other calcium-based compounds that will scale RO membranes. These compounds are calcium sulfate, phosphate, and fluoride. Although there are no specified feed water guidelines for these compounds, they are worth investigating.

- Calcium sulfate is a sparingly-soluble salt. As with barium and strontium sulfate, the potential to scale with calcium sulfate is high when the ion product exceeds 80% of the solubility constant. Antiscalants or sodium

Table 7.2 Solubility of calcium phosphate compounds.

Compound	Formula	pK_{sp}
Calcium Phosphate	$Ca_3(PO_4)_2$	28.9
Brushite	$CaHPO_4 \cdot 2H_2O$	6.68
Octacalcium Phosphate	$Ca_8H(PO_4)_3 \cdot 3H_2O$	49.6
Hydroxyapatite	$Ca_5(PO_4)_3OH$	57.74
Fluoroapatite	$Ca_5(PO_4)_3F$	60
Calcium Carbonate	$CaCO_3$	8.42
Barium Sulfate	$BaSO_4$	9.97

softening to remove calcium can be used to control calcium sulfate scale.

- Calcium phosphate has become a common problem with the increase in treatment of municipal wastewater for reuse. Surface waters can also contain phosphate. Calcium phosphate compounds can contain hydroxyl, chloride, fluoride, aluminum, and/or iron. Several calcium phosphate compounds have low solubility, as shown in Table 7.2. Solubility for calcium carbonate and barium sulfate are also shown by comparison. The potential for scaling RO membranes with the calcium phosphate compounds listed in Table 7.2 is high and will occur when the ion product exceeds the solubility constant. This can occur at orthophosphate concentrations as low as 0.5 ppm. Sodium softening or antiscalants *together* with low pH help to control phosphate-based scaling.
- Calcium fluoride scale can form when the concentration of fluoride is as low as 0.1 ppm if the concentration of calcium is high. Scaling will occur when the ion product exceeds the solubility constant. Antiscalants or sodium softening can be used to control calcium fluoride scale.

Calcium has also been shown to affect the deposition of natural organic matter (NOM).¹¹ Work by Schafer et. al., demonstrated that NOM in the form of humic substances deposit preferentially

on hydrophilic membranes, such as polyamide-based membranes.¹¹ The presence of calcium resulted in high flux decline due to precipitation of primarily humic acids (due to their relatively low molecular weight and hence lower diffusion away from the membrane in the concentration polarization boundary layer). The higher the calcium concentration, the faster the flux declined.¹² Calcium binds to the acidic functional groups of NOM resulting in a compact fouling layer on the membrane surface. Bridging between deposited NOM molecules is enhanced in the presence of calcium, leading to additional compactness of the fouling layer.¹² Operation at low flux, low trans-membrane pressure, and high shear was shown to reduce the deposition of NOM on membrane surfaces and, therefore, fouling of the membrane.^(11,12)

As described previously, calcium can be removed or reduced in RO feed water using sodium softening or lime softening (see Chapters 8.1.6 and 8.3, respectively).

7.12 Exposure to Other Chemicals

Exposure of a thin-film composite membrane to a variety of organic compounds can result in swelling or dissolution of the polysulfone microporous support layer.¹³ Suspect chemicals include:

- Solvents: dimethyl formamide, dimethyl acdimide, n-methyl pyrrolidone, dimethyl sulfoxide, etc.
- Aromatic compounds: benzene, toluene, xylene, phenol, diesel fuel, gasoline
- Others: ketones, aldehydes, esters, strong ethers

Note that only low-molecular solvents such as alcohols (isopropanol and smaller) are acceptable.

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Techniques and Technologies

Adequate pretreatment is one of the fundamental keys to successful and cost-effective operation of an RO system. Pretreatment is designed to prevent or minimize membrane fouling, scaling and degradation of membrane performance and materials. This chapter covers mechanical and chemical techniques and technologies that are commonly used to pretreat RO systems.

Figure 8.1 Shows the projected performance of an RO membrane system with ideal, marginal and inadequate pretreatment.¹ After an initial period over which time new membranes stabilize performance, a system with ideal performance will show only a slight decline in performance with time due to compaction and the inevitable fouling and scaling that will occur despite good pretreatment and system hydraulics. Marginal pretreatment exhibits more rapid decline in performance than the system with ideal pretreatment. Initial cleaning may be able to revive most of the performance, but after time, foulants and scale that were not removed become irreversibly attached to the membrane and cannot be cleaned away. The RO system with inadequate pretreatment will show very rapid decline in performance that typically cannot be recovered by cleaning the membranes. An RO system with less than ideal pretreatment faces frequent cleaning intervals and short membrane life. Frequent cleaning and membrane replacement costs money, time, and the environment.

Once optimized, the pretreatment system must be continuously evaluated and re-optimized to adjust to changes in performance of each unit operation, due to mechanical difficulties or changes in the influent water quality.

Pretreatment techniques and technologies can be categorized into four (4) general types:

1. Mechanical
2. Chemical
3. Mechanical plus chemical
4. Sequenced

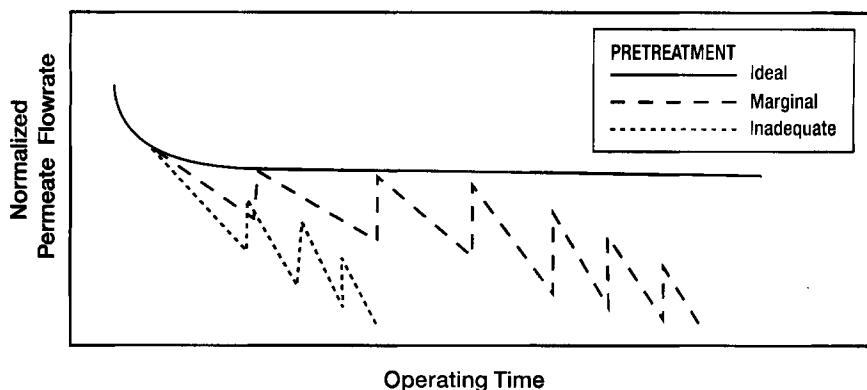


Figure 8.1 Projected performance of an RO system as a function of the quality of feed water pretreatment.

Appropriate pretreatment techniques and/or technologies for a given RO application needs to be selected based on the quality of the influent water to be treated by RO. Some water, such as well water with low concentrations of iron and manganese, may require very little, if any, pretreatment, while other water, such as river or lake water, may require extensive pretreatment using sequenced techniques and technologies.

Obtaining historical influent water quality data as well as pilot testing of proposed pretreatment unit operations are both good practice in designing and optimizing the pretreatment system.

8.1 Mechanical Pretreatment

Mechanical pretreatment involves physical techniques to reduce turbidity, suspended solids, SDI, bacteria, hardness, and heavy metals present in RO influent water. Table 8.1 lists some mechanical treatments and what species they will treat. It is important to reduce or eliminate these species from RO influent water to minimize fouling and scaling of the membranes.

8.1.1 Clarifiers

Clarifiers are used to remove large suspended solids, colloids, organics, and color from surface water supplies. Coagulation and flocculation using chemical treatments, and sedimentation or “settling” are the three primary steps used to achieve reduction of contaminants. However, the typical effluent quality from a clarifier is not low enough

Table 8.1 Mechanical RO pretreatment techniques and species that these techniques address.

Mechanical Pretreatment	Species Addressed
Clarification	Suspended solids, Colloids, Organics, Color, SDI
Multimedia Filtration	Turbidity, Suspended solids down to 2 – 10 microns, SDI
High-Efficiency Filtration	Suspended solids down to 0.25 microns
Carbon Filters	Total organic carbon, Chlorine
Iron Filters	Iron, Manganese, Hydrogen sulfide
Sodium Softeners	Hardness, Soluble iron
UV Radiation	Organics, Microbes
Membrane	Microbes, Algae, Color

in turbidity and suspended solids to send directly to an RO. Clarification is a good bulk removal technology for reducing the majority of suspended solids and turbidity. However, multimedia filtration is generally required to polish clarifier effluent to reduce the turbidity (and SDI) low enough to meet RO influent standards. See Chapter 7 for more information about RO influent water specifications.

There are three basic designs for clarifiers (also known as “gravity clarifiers”): solids-contact, inclined-plate settlers, and sedimentation. The advantages of each type of unit are:

- Solids contact: lowest chemical demand and higher effluent quality
- Inclined-plate: smallest footprint
- Sedimentation: least sensitive to fluctuations in influent flow rates—typically used for wastewater treatment.

All designs share some common design characteristics:²

- Rise Rate: Rise rate of water is the flow rate (total hydraulic load) divided by the surface area of the clarifier. Most clarifiers operate between 0.75 gpm/ft² and 1.25 gpm/ft², with the exceptions of inclined-plate

settlers, which operate at higher rise rates, as high as 2.0 gpm/ft² and the Actiflo® recirculation clarifier which operates at a rise rate of at least 16 gpm/ft² (Actiflo is a registered trademark of Veolia Water North America, Chicago, IL).

- **Rapid Mix Zone:** This is the area of the clarifier (or just immediately prior to the clarifier) where coagulation takes place.
- **Slow Mix Zone:** This is the area of the clarifier where flocculation takes place.
- **Rake:** The purpose of the rake is twofold: direct the settled solids to the blowdown and to control the character of the sludge bed. High rake speed results in a fluffier bed while low speed hinders the contact between newly-formed floc (agglomerated suspended solids) and the bed. Note that most inclined-plate clarifiers do not have this feature.
- **Sludge Bed:** The sludge bed provides filtering for floc, thereby improving the effluent quality. The height of the bed depends on the specific clarifier design, the type of chemical treatment program used, and the operating protocol.

8.1.1.1 *Solids-Contact Clarifiers*

Solids-contact clarifiers (also known as “upflow” clarifiers) typically have four treatment zones:

- **Rapid mix zone:** sludge is recirculated to this zone to improve coagulation,
- **Slow mix zone:** this is the zone where particle flocculate and settle,
- **Sedimentation zone:** the sludge bed forms in this zone,
- **Clarifier water zone:** this zone is where the clarified water exits the unit.

Solids-contact clarifiers may be circular or rectangular in design. Typical rise rates for circular clarifiers range from about 0.75 – 1.25 gpm/ft²; residence time is about one to two hours. In the rectangular Actiflo® clarifier, rise rate is significantly higher, as described above, which yields a higher throughput for this type of clarifier. As a result,

the residence time is shorter and the footprint is significantly smaller for an Actiflo® than that of a conventional circular clarifier.

Solids-contact clarifiers are characterized as either recirculation or sludge blanket clarifiers (Figures 8.2, 8.3 and 8.4). In a circular recirculation clarifier, high floc volume is maintained by recirculation from the slow mix zone to the rapid mix zone. In the rectangular Actiflo® recirculation clarifier, microsand is recirculated through the unit to provide a ballast for flocs formed during the coagulation stage. The sand acts as a seed for floc formation and provides weight to speed settling of the floc. The clarification tank is fitted with lamella to speed the settling of the microsand-ballasted sludge (see Chapter 8.1.1.2 for more discussion about lamella used in clarifiers). Finally, with a sludge blanket clarifier, high floc solids are maintained by flowing water through a fluidized blanket of solids.

8.1.1.2 Inclined-Plate Clarifiers

Inclined plate clarifiers, also known as Lamella® clarifiers, use several inclined plates (or sometimes tubes) to maximize the settling area for a given floor area (Lamella is a registered trademark of Parkson Corporation, Ft. Lauderdale, FL). Figure 8.5 shows a diagram of an inclined-plate clarifier.

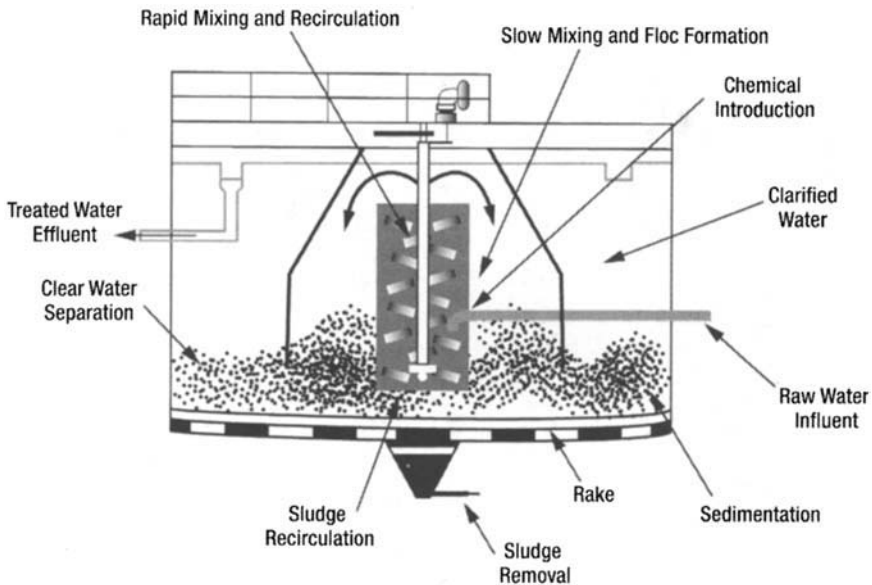


Figure 8.2 Recirculation solids-contact clarifier. Courtesy of Ecodyne Limited.

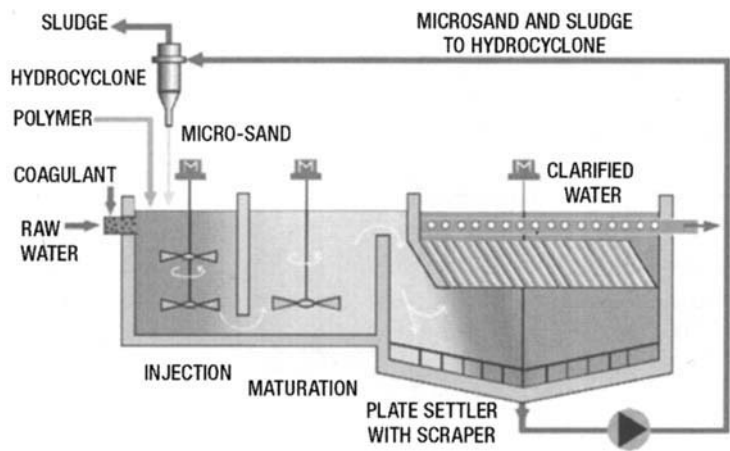


Figure 8.3 Actiflo® rectangular recirculation clarifier. *Courtesy and registered trademark of Veolia Water North America, Chicago, IL.*

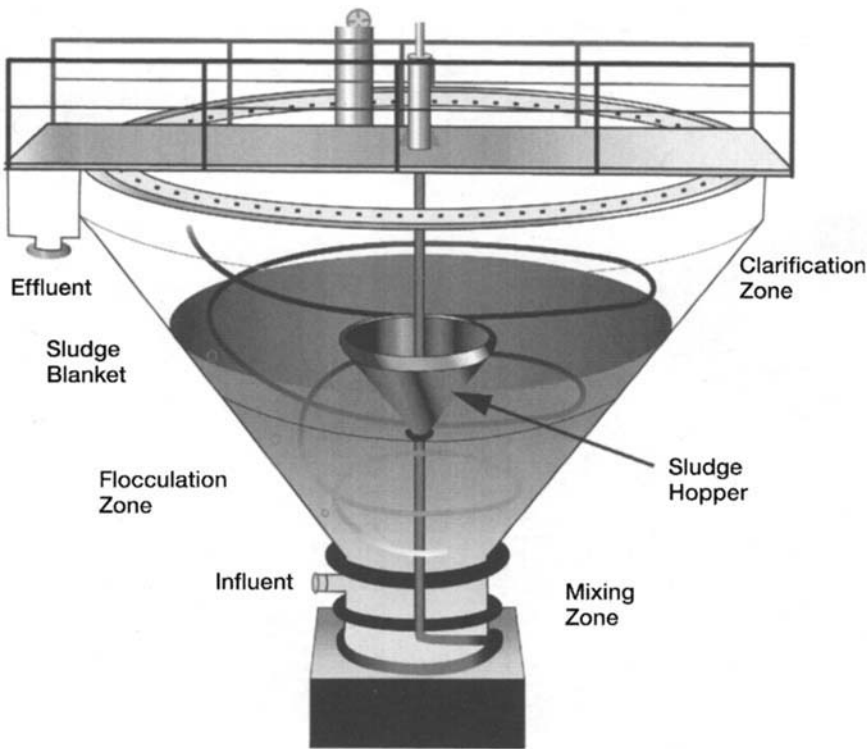


Figure 8.4 Spiracone® sludge-blanket clarifier. *Courtesy and registered trademark of Siemens Water Technologies Corp., Warrendale, PA.*

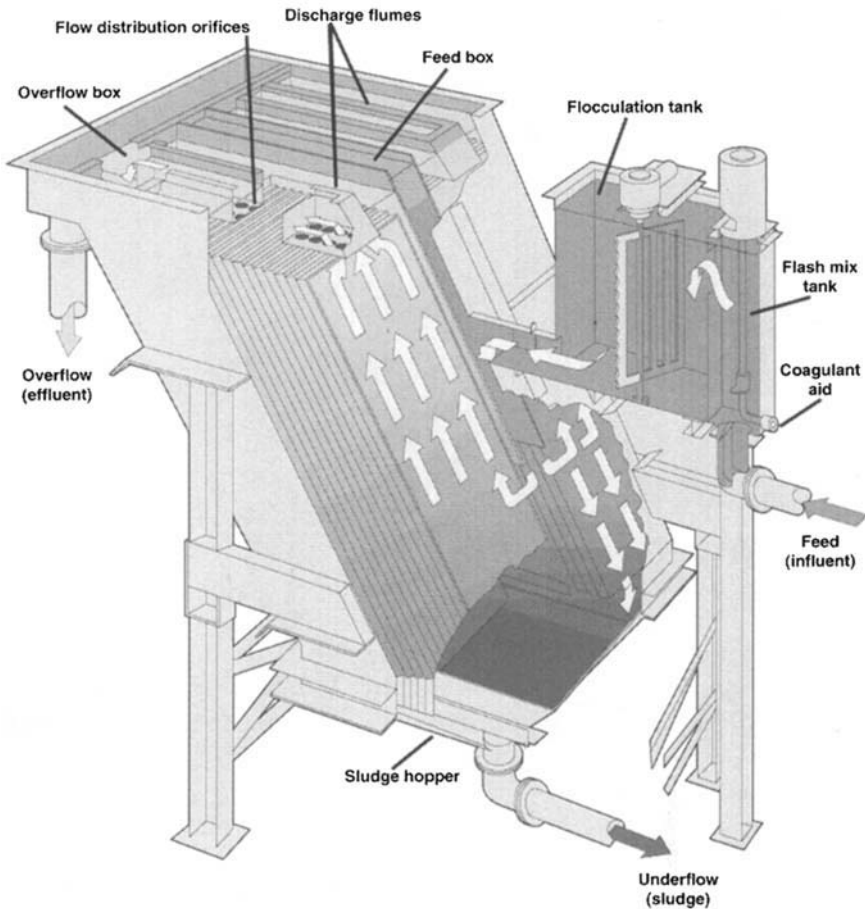


Figure 8.5 Inclined-plate clarifier. *Courtesy of Parkson Corporation.*

Inclined-plate clarifiers have five treatment zones:

- Flash mix tank: coagulant is feed in this rapid-mix zone
- Flocculation tank: flocculants are fed in this tank
- Settling zone: plates (or tubes) on which the solids settle,
- Outlet (Overflow) zone: this is where clarified water exits the unit,
- Sludge (Underflow) zone: area where sludge collects and exits the unit

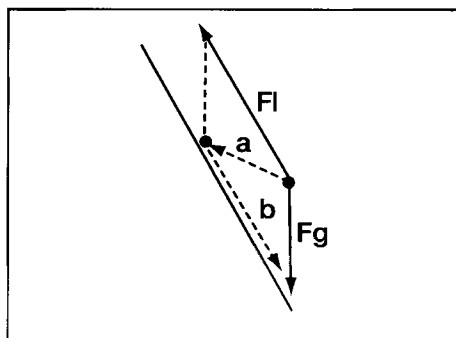


Figure 8.6 Particle motion in a lamella clarifier, where F_l is the convective flow force vector and F_g is the gravitational force vector. Particles first move toward the lamella (a) and then move toward the sludge zone (b).

Rise rate is about 2.0 gpm/ft² in an inclined plate clarifier.

The plates or “lamella” collect solids. The incline of the plates is a function of the density of the solids; typical set point is about 55° from horizontal. The settling distance varies from a few inches to a maximum of a few feet, unlike conventional clarifiers, where the settling distance can be several feet. Figure 8.6 shows how the particles between the lamella plates migrate to the plate surface (a) following the resultant vector of the fluid drag (F_l) and gravity (F_g) forces. Once on the plates, the particles slide down to the sludge zone (b).

The advantages of this arrangement over conventional, circular clarifiers include:

1. the footprint of the system is much smaller
2. the plates minimize sludge carry over
3. the unit is ready to start up and stop at any time without operational delays
4. the unit provides high efficiency separation with minimum-density floc
5. the unit exhibits, good performance in low-turbidity water, with dirty-water recycle

8.1.1.3 Sedimentation Clarifiers

Sedimentation clarifiers are generally used for secondary clarification of wastewater. They are characterized by having four treatment sections:

- Inlet zone: coagulant is fed into the influent of this mixing zone,

- Settling zone: this is the zone where particles settle,
- Outlet zone: Weirs direct effluent out of the clarifier,
- Sludge zone: This zone is where solids settle, and are collected for removal.

Sedimentation clarifiers may be circular or rectangular in configuration. In a circular sedimentation clarifier, the influent enters into the center or the perimeter (“peripheral feed”) of the unit. In a centered unit, the effluent water is collected in the perimeter. It is difficult to control the recirculation in a perimeter-fed unit, and hence, is not common. In a rectangular sedimentation unit, the flow is linear. Rise rate is about 1 gpm/ft² with a residence time of two to six hours.

8.1.1.4 Chemical Treatment for Clarifiers

Chemical addition is typically used with clarification to improve both the utility and performance of the unit operation. Coagulants and flocculants are generally used to improve the ability to settle particles in the clarifier. Jar tests are used to determine the proper dosage of chemicals and streaming current detectors or turbidity monitors are used to monitor performance and control chemical dosage. Chlorine is often used to improve the removal of organics and color in the clarifier. Chlorine also provides disinfection of the make-up water to prevent the clarifiers from going septic.

8.1.1.4.1 Coagulation

Coagulation involves using cationic compounds to charge neutralize the net negative charge of suspended particles in water. Charge neutralization of the net negative charge on particles allows the particles to move closer together in anticipation of creating larger particles that settle faster. Coagulation requires rapid mixing and occurs immediately upstream or in the influent well of the clarifier, depending on design.

Coagulants can be inorganic salts or organic compounds. Inorganic salts include aluminum sulfate (alum), ferric sulfate, and ferric chloride. Organic polymers used for coagulation such as polyamines and poly(diallyl-dimethyl-ammonium chloride), commonly known as poly-DADMACS, are generally lower-molecular weight (<500,000), high-charge cationic polymers.

The performance of inorganic coagulants depends on pH. The performance of alum is optimum at pH 6.0 – 7.0 (with an optimum at pH = 6), while ferric coagulants can be used over a broader pH range, from 5.0 – 11.0 (with an optimum at pH = 8).

The performance of polymeric coagulants depends on the amount of turbidity present in the water. At less than 10 NTU, an inorganic or combination inorganic/organic polymer is preferred. At 10–60 NTU, a combination of inorganic and organic coagulants are generally used. At greater than 60 NTU, a polymeric coagulant alone is sufficient.

Caution must be used whenever coagulants are used upstream of an RO due to the potential for fouling the membrane with the coagulant or its constituents. Alum and ferric carryover is common and the aluminum and iron will foul the RO membranes. Under conditions of poor pH control or high surface water turbidity, high doses of alum can result in post precipitation of aluminum. Aluminum hydroxide is difficult to remove through filtration and will foul RO membranes. Operators of RO systems that are located downstream of an alum feed need to be extra vigilant in monitoring membrane performance for signs of fouling. Membranes need to be cleaned when normalized product flow or pressure drop data indicate that performance has dropped by 10% to 15% from start-up (see Chapter 13.2.1).

Cationic polymeric coagulants have a natural affinity for the negatively-charged polyamide composite membranes, leading to irreversible fouling of the membrane. Therefore, overfeed and carryover of polymeric cationic coagulants must be carefully monitored and avoided. Additionally, cationic polymers can co-precipitate with negatively-charged antiscalants and further foul an RO membrane.

8.1.1.4.2 Flocculation

Flocculation is the agglomeration of charge-neutralized particles into larger particles. Unlike coagulation, rapid mixing is not required for flocculation to take place. It typically occurs in the reaction chamber or “slow mix” zone of the clarifier.

In general, very high molecular-weight, anionic polyacrylamides are the most effective flocculants. Some flocculation success may occur using non-ionic polymers. Typical anionic flocculants are 1,000,000 to 2,000,000 molecular weight, while non-ionics can be up to 20,000,000 molecular weight. Flocculants, due to their anionic or non-ionic nature, do not cause the same degree of fouling of negatively-charged polyamide composite membranes that cationic coagulants do, but overfeed can lead to organic fouling of the membranes by the flocculant polymer that can usually be removed if addressed in a timely manner.

8.1.1.4.3 Chlorine

Chlorine is usually added upstream of a clarifier to oxidize organics, to improve the removal of color in the clarifier, and to control

microbial growth in the clarifier and downstream equipment. Chlorine along with an alum feed at pH 4.5 to 5.5 is optimum for color removal. This is important for RO pretreatment, as color can irreversibly foul a polyamide composite membrane (see Chapter 8.2.1.1 for a more detailed discussion about chlorine for RO pretreatment).

8.1.2 Multimedia Pressure Filters

Multimedia pressure filters are designed to reduce turbidity and colloids (measured as SDI) in water. These filters can remove particles down to about 10 microns in size. If a coagulant is added to the filter influent stream, reduction of particles down to 1-2 microns can sometimes be accomplished. Typical removal efficiency for multimedia pressure filters is about 50% of particles in the 10 – 15 micron size range. Influent turbidity for RO pretreatment is limited to about 10 NTU. At turbidity greater than 10 NTU, these filters may backwash too frequently to provide consistent effluent quality at reasonable run lengths.

Multimedia pressure filters contain graduated layers of anthracite on top of sand on top of garnet. Figure 8.7 shows a cross section of a multimedia filter. The fine garnet material is denser than the coarse anthracite material. There is no discrete boundary between each of the layers; there is a gradual transition from one material density and coarseness to the next. Otherwise, there would be a build up of particles at each interface. Particles are subsequently removed through the filter using physical entrapment. Larger particles are removed on top through the anthracite, while smaller particles are subsequently removed through the sand and garnet. Multimedia filters offer finer filtration than dual media (anthracite and sand) filters due to the relatively fine nature of the garnet.

Service flow rates for RO pretreatment should be about 5 gpm/ft² of media. Throughput can be estimated using a filter about 0.45 lb of suspended solids per square foot of filter loading of area. Backwash rates should be 15 gpm/ft² at 60°F. Lower water temperatures require higher flow rates to adequately raise the bed for a complete backwash. A 30 – 50% bed expansion is necessary to achieve good backwashing of the media. Raw, unfiltered water can be used for backwash. Alternatively, a filtered product tank can be provided that also acts as a reservoir for backwash water. Some operators choose to recovery RO reject by using this water to backwash filters. However, for best results, use of RO reject water to backwash a multimedia filter is not recommended. In some applications where

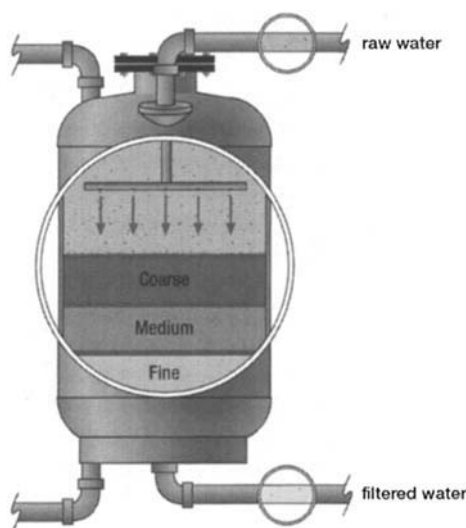


Figure 8.7 Multimedia pressure filter showing coarse, medium and fine media, typically anthracite, sand, and garnet, respectively.

there is heavy use of coagulants, an air scour during backwash is recommended. An air scour involves air that is bubbled up through the bed to help release some stubborn solids from the media.

Multimedia pressure filters operate best when they are continuously on line. When multiple filters are required, all filters will be on line except during backwash, when only one filter is off line for backwashing. During backwash of one filter, the remaining filters on line will handle the full influent flow rate. To achieve the 5 gpm/ft² design flux as an average service flow rate, the diameter of the filters should be such that when all filters are on line, the flux is slightly less than 5 gpm/ft², and during backwash of one filter, the flux through the remaining filters on line should be less than about 7.5 gpm/ft².

Multimedia pressure filters can be vertically or horizontally oriented. Figure 8.7 shows a vertical filter while Figure 8.8 shows a horizontal multimedia pressure filter. Horizontal multimedia filters are separated internally into “tanks” or compartments. Each tank acts as an individual filter. When it is time to backwash one of the tanks, the effluent from the other tanks provides the backwash water. The key in selecting horizontal filters is that the filter should have enough internal tanks so that productivity (required effluent flow rate plus the

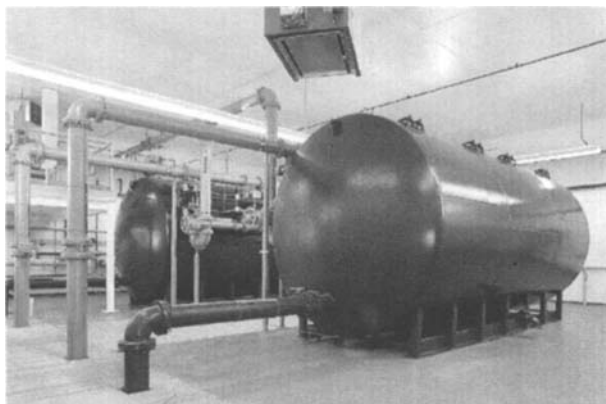


Figure 8.8 Horizontal multimedia pressure filter with 4 internal tanks.
Courtesy of Siemens Water Technologies Corp.

flow rate needed to backwash one tank) can be maintained even when one tank is in backwash. Given the design of a horizontal filter, effluent from all tanks is combined and a portion of this total becomes backwash water for one tank. Horizontal filters are used at higher flow rates, as the footprint for the horizontal filters is smaller than that for vertical filters for the same throughput.

Multimedia pressure filters can be used as stand alone treatment, when the feed sources is relatively clean water, such as a potable municipal supply, or in series with clarification, for river and other surface waters.

Multimedia pressure filters can be used in conjunction with filter aids (typically polymeric coagulants) to increase the removal efficiency of the media. In essence, the coagulant “bridges” particles to allow them to agglomerate so that they are more easily removed by the filter media. There is an optimum dosage of filter aid at which the turbidity removal is optimum. Too much or too little filter aid will decrease the performance, as shown in Figure 8.9. The optimum dosage is found empirically for each application, with ranges from about 3 – 10 ppm for inorganic coagulants and 0.25 – 2.0 ppm for organic and inorganic/organic blends.

8.1.3 High-Efficiency Filters

High-efficiency filters (HEF) are pressure filters designed to remove turbidity and SDI in water. They were originally applied to cooling towers, to reduce the concentration of suspended solids in those systems.

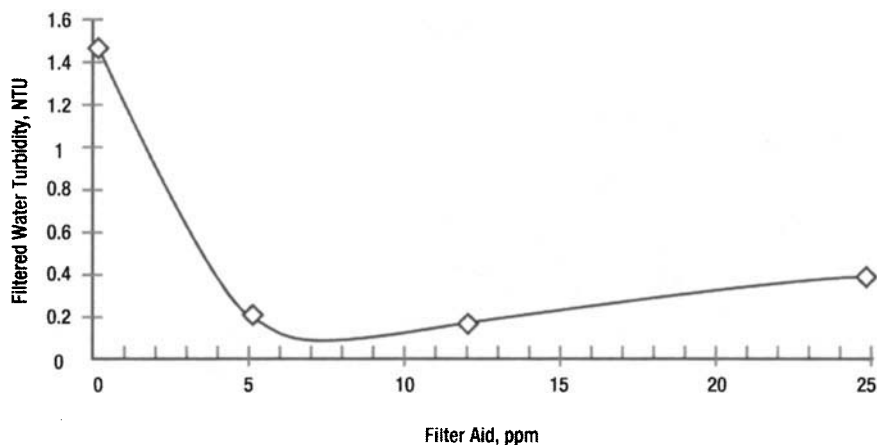


Figure 8.9 Effect of filter aid dosage on effluent turbidity from a multimedia pressure filter. Too little cationic coagulant feed does not charge neutralize the net negative charge on the particles thereby not reducing turbidity enough. Too much cationic coagulant changes the charge surrounding the particles from a net negative to a net positive, resulting in charge repulsion and higher effluent turbidity.

Table 8.2 shows a typical cooling water particle size distribution. As seen in the table, nearly 98% of all particles are smaller than 0.5 microns, and greater than 99% of all particles are smaller than 5 microns. Recall that multimedia filters were only about 50% efficient at 10–15 microns. By comparison, HEFs can remove 50% of particles as small as 0.25 microns in size.

There are two basic HEF designs, the top-over-bottom design and the vortex design. Figures 8.10 and 8.11 show cross sections of each of these types of HEFs.

Vortex filters use centrifugal force to swirl raw water above the surface of the media in a cross-flow manner. Large suspended solids are collected on the sidewall of the filter tank. The smaller solids drop down to the surface of the fine sand media and are filtered out through this media. When the pressure drop reaches about 15 psig, the vortex filters are backwashed.

Vortex filters operate at high flow rates, typically 15–20 gpm/ft². Backwash flow rate is about 2.5–5 gpm/ft² and requires only 4–8 minutes of backwash time.

Vortex filters supplied by Sonitec, Inc., (“Vortisand®” filters—Vortisand is a registered trademark of Sonitec, Inc., Holyoke, MA) come complete with a chemical cleaning system. This system is used every 3–8 backwash sequences to reduce filter media fouling

Table 8.2 Typical cooling water particle size distribution.

Size (micron)	Particle Count		
	Per 100ml	Percent of Particles in this Size	Cumulative Percent of all Particles
0.5 – < 1.0	30,277,895	97.7	97.7
1.0 – < 5.0	614,664	2.0	99.7
5.0 – < 10	72,178	0.2	99.9
10 – < 15	21,561	0.1	100.0
15 – < 20	8,186	0.0	100.0
>20	5,765	0.0	100.0
Total	31,000,249	100.0	100.0

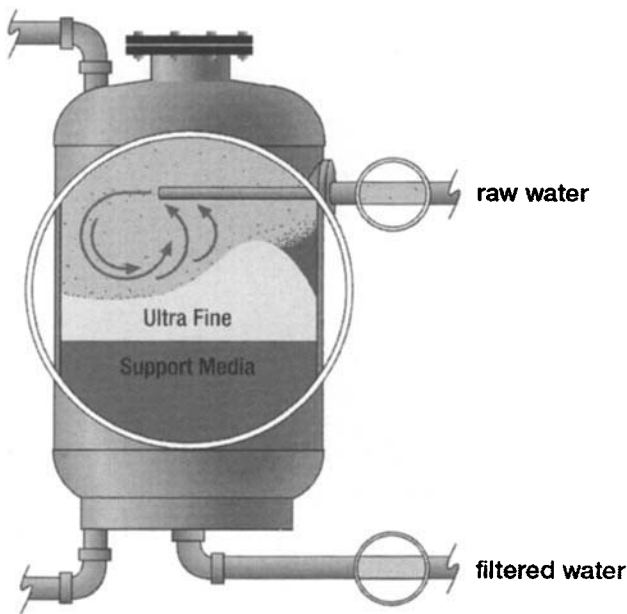


Figure 8.10 Top-over-bottom high efficiency filter.

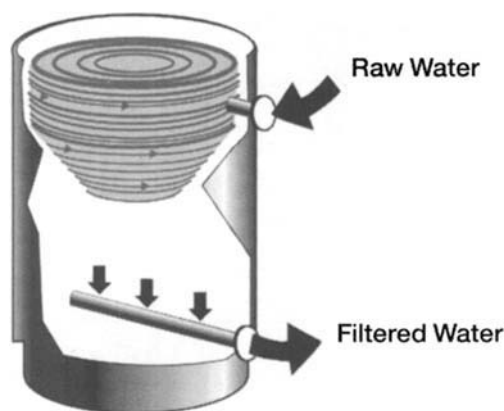


Figure 8.11 Vortex high efficiency filter. *Courtesy of Sonitec, Inc.*

and maintain filtration effectiveness. Dispersants are typically used in the spring when high run off can result in fouling of the filter media with silt. Chlorine is necessary during the summer and fall to reduce biofouling.

Top-over-bottom HEFs rely on tangential flow across the surface of the media to affect filtration. Raw water enters the filter tangentially to the surface of the media, thereby creating turbulence over the bed and a tangential force that scrubs suspended solids off of the surface of the media. The tangential force also causes some of the sand in the filter bed to collect near the inlet of the filter into what looks like a “camel hump.” Solids that have been swept off of the filter surface collect behind this hump, as this area, right underneath the inlet to the filter, is an area of low turbulence. Solids continue to collect here until they spill over the hump on to the high-turbulent area of the bed. As more solids collect on the turbulent side of the filter bed, the pressure drop through the filter increases. When the pressure drop reaches about 18 psig, the filters should be backwashed. Note that the filters should be backwashed at least once per day to avoid fouling of the bed.

Top-over-bottom HEFs operate at higher flow rates, such as 10 – 15 gpm/ft². Backwash flow rate is about 10 gpm/ft² for about 5 minutes per backwash.

Table 8.3 compares performance parameters of multimedia pressure filters and HEFs. The higher throughput of HEFs reduces the footprint of the system required when compared to multimedia filters. Also, the lower backwash flow requirements for HEFs leads to less waste water to dispose of and smaller backwash components on these filters.

Table 8.3 Performance comparison of multimedia and high-efficiency filters.

Parameter	MMF	HEF
Service flow, gpm/ft ²	5	10–15
Max differential pressure, psig	15–25	18
Capacity, lb/ft ³	0.45	NA
Inlet turbidity, NTU	10*–30	1–200
Backwash flow rate, gpm/ft ²	15	10
Backwash time per backwash, minutes	30–60	5

* Maximum turbidity for RO pretreatment

N/A = not available

High-efficiency filters are gaining in acceptance for RO pretreatment.³ These filters offer some advantages over conventional multimedia filters, the most important of which may be the ability to remove particles down to 0.25 microns in size for the top-over-bottom filters, and 0.45 microns for the vortex filters.

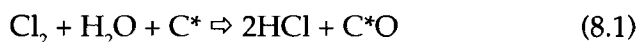
8.1.4 Carbon Filters

Activated carbon filters are used to reduce the concentration of organics in RO feed water. These filters are also used to remove oxidants such as free chlorine from RO feed water.

Activated carbon is derived from natural materials such as bituminous coal, lignite, wood, fruit pits, bones, and coconut shells, to name a few. The raw materials are fired in a low oxygen environment to create char, which is then activated by steam, carbon dioxide or oxygen. For most industrial applications, bituminous carbon is used. This is because of the smaller pores size, higher surface area and higher density than other forms of carbon giving bituminous carbon higher capacity for chlorine. Carbon can also come in one of 3 forms: powdered (PAC), extruded block (CB), and granular (GAC). Most industrial applications used GAC as this is the lowest cost of the 3 types of carbon media and this type of carbon can be reused.

All carbon is characterized by high surface area. A gram of carbon can have surface area in excess of 500m², with 1,500 m² being achievable.⁽⁴⁾ High surface area is necessary for reduction of organics and chlorine within reasonable residence time.

Chlorine and other oxidants are removed using activated carbon by an oxidation/reduction reaction. Chlorine oxidizes the carbon while the chlorine is being reduced. Chlorine ends up forming hydrochloric acid via equation 8.1.



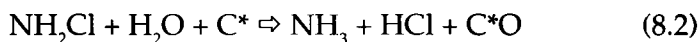
where:

C^* = carbon

C^*O = oxidized carbon

The reaction in equation 8.1 is virtually instantaneous. Effluent concentrations of chlorine are typically less than 0.05 ppm, which is adequate quality for feed to a downstream RO.

Activated carbon can also be used for chloramine removal, but the reaction time is much longer, about 5 – 10 minutes in a new bed and up to 30 minutes in a bed near equilibrium. The reaction with chloramine is a two-step process:



Note that the carbon filtration removal of chloramines leaves behind some residual ammonia. Recall from Chapter 3, Table 3.2, that ammonia as the gas is not rejected by polyamide RO membranes. The pH of the solution must be below about 7.8 for the ionized form of ammonia (ammonium) to be present and well rejected by the RO membrane. Furthermore, ammonia gas swells the RO membrane leading to lower rejection of dissolved solids by the membrane. See Chapter 8.2.1.1 for more discussion about chloramines.

The removal of organics (typically measured by total organic carbon or TOC), is an adsorption process. The surface of activated carbon is both hydrophobic and oleophilic, conditions favorable for good removal of TOC. The capacity for organics is a function of temperature, pH, nature of the organic, and concentration such that accurate capacity predictions are not possible. However, typically 25% to 80% of TOC is removed through a carbon filter.

The iodine and molasses numbers are used to determine the capacity of carbon for micro- and macro-molecules, respectively. Iodine number is a measure of the micropore (0 to 20 angstroms)

content of the carbon. It is equivalent to about 900 to 1,100 m²/g. Typical range for iodine number is 500 – 1,200 mg/g. A high number indicates a greater capacity for small molecules. The molasses number is a measure of the macropore (greater than 20 angstroms) content of the carbon. Typical range for molasses number is 95 – 600, with higher numbers indicative of a higher adsorption capacity for large molecules. Note that the European molasses number is inversely related to the North American molasses number.

Carbon filters are not designed to remove suspended solids or bacteria. In fact, carbon filters encourage the growth of bacteria, with the organics removed through the media providing nutrients for the microbes.^{4,5} This invites caution when using carbon for pretreatment prior to RO. Carbon fines are continuously sloughing off of the bed. These fines may be infected with bacteria, which can then get into the RO system and can foul the membranes. Periodic servicing of the carbon by hot water or steam sanitization is required to destroy bacteria in the bed. Note that using carbon as an oxidant removal mechanism does not eliminate or reduce the need to service the carbon unit for biofouling.

Carbon filters have the following influent water requirements to assure optimal operation of the filter:

- Turbidity < 5 NTU
- Free chlorine: <10 ppm
- TOC: <5 ppm

Service flow rates for carbon filters when used to pretreat RO feed water are as follows:

- TOC removal: 1 gpm/ft³
- Chlorine removal (pre RO): 2 gpm/ft³

The filters are backwashed occasionally to remove any suspended solids that may have accumulated on the surface of the bed. Backwashing does not remove material adsorbed in the pores of the carbon. Although a few installations regenerate their carbon using thermal, steam, solvent extraction, or other techniques, most applications see replacement of carbon when exhausted. Typical “life” of carbon used for TOC removal is 6 – 12 months. For chlorine removal, the typical “life” is 12 – 18 months.

Carbon filters were once the standard method for removing chlorine from RO influent water. However, due to the microbial fouling

that occurs downstream of a carbon filter, the current primary RO application is for the reduction of TOC only.

8.1.5 Iron Filters

Many well waters contain soluble iron, manganese, and hydrogen sulfide that oxidize in the presence of oxygen or chlorine to form insoluble hydroxides and elemental sulfur, all of which foul RO membranes (in the case of elemental sulfur, the fouling is irreversible).⁵ Manganese dioxide media is used to oxidize and filter out the oxidized metals. Specifically, manganese greensand and alternatives such as BIRM® (sometimes called better iron removal media) and Filox, are three types of media containing manganese dioxide that are used to oxidize and filter iron, manganese and the like (BIRM is a registered trademark of Clack Corporation, Windsor, Wisconsin). Table 8.4 compares properties of these three media. As the table shows, Filox contains the most manganese dioxide and has the longest life expectancy of the three media. Table 8.5 compares some additional properties of Filox, BIRM®, and manganese greensand media.

Table 8.4 Properties of manganese greensand, BIRM®, and Filox.

Parameter	BIRM®	Manganese Greensand	Filox
pH	6.5 for Iron 8 – 9 Manganese	> 6.8	5.0 – 9.0
Hydrogen Sulfide	Not Compatible	OK	OK
Turbidity	< 10 NTU	< 4 NTU	NA
Service Flow Rate, gpm/ft ²	4	5	6
Bed Depth, inches	30 – 36	24	20
Backwash Flow Rate, gpm/ft ²	12	12 – 13	12 – 15
Backwash Duration, minutes	20 – 30	15	NA
Relative Service Life	1	50	7500 (unlimited)

NA = not available

Table 8.5 Additional properties of BIRM®, manganese greensand, and Filox.

Specification	BIRM®	Manganese Greensand	Filox
Manganese Dioxide (%)	< 0.01	0.5	75 – 85
Material		Glaucanite	Filox ore
Color	Black	Black	Grey-Black
Form		Granular	Granular
Screen Size (mesh)	9 x 35	18 x 60	12 x 40
Density, lb/ft ³	44 – 50	85	114
Effluent (ppm)		0.1 or 98% reduction	0.05
TDS Limit (ppm)		850	> 1100
Capacity ppm (unless noted)			
• Iron	10 ppm	550*	27 ppm
• Manganese	5 ppm	400*	11 ppm
• Hydrogen Sulfide	0 ppm	175*	17 ppm

*grains

8.1.5.1 Manganese Greensand Filters

Manganese greensand has been used in the United States since the 1920's. Manganese greensand is a natural zeolite with an exchange capacity of about 3,000 grains/ft³. It was used extensively for softening applications until the development of synthetic gel-type resins following World War II. It is commonly known as New Jersey greensand and is derived from glauconite (an iron-potassium-silicate zeolite of marine origin). Stabilized glauconite is coated with manganese oxide in various valence states. It is the coating that provides the oxidative properties of the material.

Manganese greensand requires the use of oxidizers to aid in the oxidation and removal of iron, manganese, and hydrogen sulfite. Iron can be removed with the use of chlorine as the oxidizer. Manganese removal via manganese greensand requires the use of potassium permanganate to adequately oxidize the metal.

Manganese greensand can become exhausted and therefore, require regeneration. Regeneration can be on a continuous or intermittent basis. For continuous regeneration without manganese present, iron can be removed with 1 ppm of chlorine per ppm of iron. For iron and manganese removal, the potassium permanganate demand is about 1 ppm per 1 ppm of iron plus 2 ppm per 1 ppm of manganese (pH should be greater than about 7.5 for optimum manganese oxidation). For intermittent regeneration, the dosage of potassium permanganate should be 1.5 – 2.0 ounces per cubic foot of media.

Effluent concentrations of iron and manganese are limited to 2% of the influent concentration (98% removal) or 0.1 ppm iron, whichever is greater.

Manganese greensand filters are periodically backwashed to remove the precipitated metals. Backwashing should be initiated at 10 psig pressure drop or at a filter loading of 700 grains of iron removed per square foot of vessel area, whichever comes first (see Table 8.5 for capacity of manganese greensand).

Low concentrations of hydrogen sulfide can be treated with manganese greensand. However, the greensand will degrade at a rate proportional to the concentration of hydrogen sulfide. At some point in time, the greensand will need to be replaced due to loss of functionality.

Waters that may be contaminated with high concentrations of turbidity may want to consider a top layer of anthracite to capture these solids. This protects the greensand and ensures that the greensand is free and clear to proceed with oxidation and filtration of metals. A 15 – 18-inch layer of anthracite is recommended.

Because of the popularity of manganese greensand, occasional shortages and long lead times have been experienced. Hence, the development of alternatives to manganese greensand. These alternatives include BIRM® and Filox. In general, these and other alternatives to manganese greensand exhibit higher capacity, efficiency, and life. Alternatives to manganese greensand are discussed below.

8.1.5.2 BIRM® Filters

BIRM® or “better iron removal media” filters are used to oxidize and filter out iron and manganese. BIRM® is a man-made granular zeolite coated with a fine dusting of manganese dioxide. BIRM® acts as a catalyst and uses the dissolved oxygen in the water to affect the oxidation of iron and manganese. The dissolved oxygen content in the water must be at least 15% of the iron content. Chlorine is not

recommended as this may deplete the coating on the media. BIRM® is not compatible with hydrogen sulfide.

8.1.5.3 *Filox Filters*

Filox is a naturally-occurring ore used for the removal of iron, manganese, and hydrogen sulfide from water. Filox contains 80% manganese dioxide in a cluster format that enhances performance, including its capacity. Filox is capable of removing up to 15 ppm of iron, 7 ppm of hydrogen sulfide, and 3 ppm of manganese *without* additional oxidizers. However, it is recommended that chemical regeneration be considered for most applications.

The need for oxidants can be determined by measuring the oxidation reduction potential (ORP) of the water to be treated. If the ORP measures above negative 170 millivolts, Filox can be used without the use of additional oxidants. Lower than negative 170 millivolts will require additional oxidants. Air, hypochlorite, hydrogen peroxide, ozone, and potassium permanganate are all suitable oxidants to use with Filox. Note that weaker oxidants, such as air and hypochlorite will be sufficient for most applications.

Effluent concentrations of iron and manganese from a Filox filter are about 0.05 ppm.

Backwashing of the Filox media is critical to successful operations. Inadequate backwashing will lead to bed fouling and eventual failure of the bed. About 20% to 50% bed expansion is required to ensure adequate backwashing. Filox is a heavy medium at 114 lb/ft³. Hence, a valve capable of allowing the lifting of the bed about 20% to 50% at 12 to 15 gpm/ft² (at 60°F) is required for successful backwashing of the media.

8.1.5.4 *Other Iron Removal Media*

There are several other media available in the marketplace, including AD26 (AdEdge Technologies), LayneOx (Layne Christensen), and DMI65 (Itochu Chemicals America), among others, that purport to be effective for iron, manganese, and even arsenic and hydrogen sulfide removal. The performance characteristics of these media vary, and they have operating characteristics that also vary. Some require more oxidizers than others. Most media are very heavy and require significant back wash flow rates and/or air scour to clear metal oxides from the bed. The best recommendation is to pilot the media of interest prior to installing a full-scale system. Also, when considering the full-scale system, the designers needs to be aware of the required

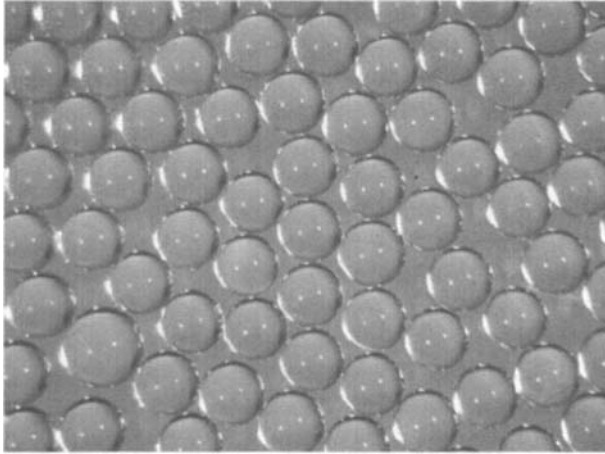


Figure 8.12 Dowex™ Upcore™ Monosphere C-600, styrene-divinylbenzene gel cation resin (Dowex and Upcore are trademarks of Dow Chemical Company, Midland, MI).

backwash flow rate which can be significant and determine whether this flow is available at the site.

8.1.6 Sodium Softeners

Sodium softeners are used to treat RO influent water to remove soluble hardness (calcium, magnesium, barium, and strontium) that can form scale on RO membranes. Once known as sodium zeolite softeners, zeolites have been replaced with synthetic plastic resin beads. For sodium softeners, these resin beads are strongly acidic cation (SAC) polystyrene resin in the sodium form. The active group is benzene sulfonic acid, in the sodium, not free acid, form. Figure 8.12 shows styrene-divinylbenzene gel cation resin.

Equation 8.4 shows the softening reaction for calcium exchange:

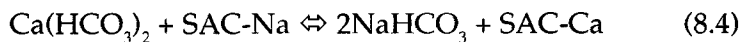
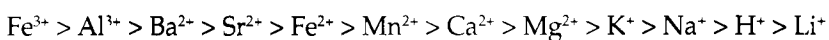


Figure 8.13 illustrates equation 8.4.(6) Calcium exchanges with sodium because the resin has selectivity for calcium over sodium. The relative selectivity of SAC resin is shown below:



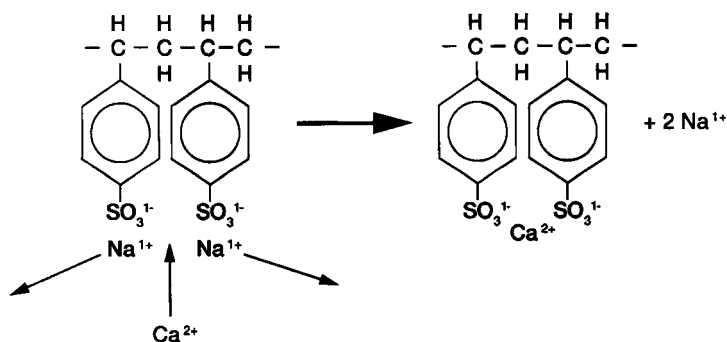


Figure 8.13 Sodium softening reaction for calcium exchange.

Assuming the selectivity of the resin for sodium is 1, the relative selectivity for magnesium, calcium, and strontium would be 1.7, 2.6, and 3.3, respectively. Selectivity is related to valence or charge on the ion. The higher the valence, the higher the SAC resin selectivity for that ion.

Note that the resin shows a preference for iron and manganese over hardness. Although sodium softeners will remove these metals, they also are not removed during regeneration. Therefore, sodium softeners operating on well waters with high iron or manganese must be periodically cleaned. The same is true of aluminum, when alum is used ahead of sodium softeners. Cleaning the resin of iron or aluminum typically involves the use of an acid such as hydrochloric, phosphoric, or citric. Following cleaning with acid, a double regeneration with salt is required to return all sites back to the sodium form (otherwise, a low pH can result during the first service run after a cleaning).

Theoretically, hardness in the effluent from a sodium softener can be less than 1.0 ppm as CaCO_3 . In practice, effluent quality from a sodium softener is dependent on the influent water quality. Higher influent hardness leads to higher effluent hardness. The total dissolved solids (TDS) concentration in the effluent from the softeners should be equivalent to that of the influent water since no TDS is removed through the softener; instead sodium replaces hardness in the treated stream. The effluent will have a much higher concentration of sodium and lower concentration of hardness than the influent has.

Effluent hardness is nearly constant for most of the service run. As the resin nears exhaustion, the hardness in the effluent begins to increase. At this point, it is time to regenerate the resin.

Regeneration of the resin involves replacing the hardness ions that have exchanged onto the resin with sodium. This is possible for two

Table 8.6 Hardness leakage and resin capacity as a function of regenerant salt dosage.

Salt Dosage, lb/ft ³	Theoretical Capacity (grains)	Hardness Leakage (ppm)		
		100 ppm total electrolyte	250 ppm total electrolyte	500 ppm total electrolyte
6	20,000	0.1 – 0.3	0.5 – 2	3 – 6
10	24,000	0.1	0.3 – 0.5	1 – 3
15	30,000	0.1	0.1 – 0.2	0.5 – 1

reasons: first, exchange Equation 8.4 is reversible and second, an excess of sodium will drive Equation 8.4 in the reverse direction. A 10% sodium chloride solution is used to regenerate resin. The dosage of salt used during regeneration will determine the capacity of the resin for hardness. Table 8.6 lists the theoretical capacity of the resin for hardness as a function of salt dosage used in regeneration. The higher the salt dosage, the higher the resin capacity is for hardness. Maximum salt dosage used in industrial applications is 15 lb/ft³ of resin. This is because the curve of resin capacity as a function of salt dosage is asymptotic; the curve tends to flatten out at higher salt dosages resulting in little increase in capacity as the salt dosage is dramatically increased.

Regeneration of softener resin is a 4-step process.

1. **Backwashing:** Backwashing the resin removes suspended solids and resin fines that may have collected in the vessel. Typical flow rates for backwashing range from about 4 to 8 gpm/ft² for a minimum of 10 minutes or until the backwash water runs clear. The target is to expand the bed by 50% for adequate backwash. Backwash water is usually sent to drain.
2. **Brining:** This step involves the injection of brine into the resin bed. A brine maker or day tank is used to dissolve salt into a 25% solution. This solution is then diluted in line with service water to 10% and either pumped or educted into the resin bed. Flow rates range from 0.5 to 1.0 gpm/ft³. The low flow rate is nec-

essary to give ions time to diffuse into and out of the resin beads. Duration of backwash is typically 20 to 30 minutes. Effluent is sent to drain.

3. Slow rinse: A slow or “displacement” rinse is used to remove traces of brine from the bed. Dilution water from the brining step continues to run at the same total flow rate as the brining step. Duration of the low rinse step ranges from 8 to 25 minutes. The water is sent to drain.
4. Fast Rinse: The fast rinse is conducted using service water at the service flow rate, typically 6 to 8 gpm/ft². The objective is to rinse out traces of brine and hardness. Duration of the fast rinse is 15 to 20 minutes or until effluent hardness returns to nominal concentration. The fast rinse is usually sent to drain.

For optimal operating of the sodium softener, the feed water to the system should adhere to these guidelines:

- Temperature: 45°F – 250°F
- Total dissolved solids: < 750 ppm
- Total hardness: < 350 ppm as CaCO₃
- TOC: < 2 ppm
- Color: < 5 APHA
- Turbidity: < 6 NTU
- Iron: < 0.2 ppm
- Manganese: < 0.1 ppm
- Free chlorine: < 0.5 ppm
- Service flow rate: 8 gpm/ft² nominal, up to 25 gpm/ft² for some polishing applications

Softening water prior to RO helps to minimize the potential for scaling the membranes with hardness. However, more and more facilities are being faced with chloride-discharge limitations, making the use of sodium softeners prior to RO undesirable. To achieve hardness-free water while minimizing the chloride discharge, sodium softeners are being used as polishers for RO effluent. See Chapter 15.1.1 for a discussion on the merits of pre- and post-RO softening.

8.1.7 Spent Resin Filters

Spent or exhausted resin has been used on occasion to filter RO influent water. These filters are designed to remove silt and reduce SDI from surface water sources. All evidence to the effect of such

filters on the suspended solids in RO influent is anecdotal; there appears to be no data in the literature attesting to the efficacy of such filters for RO influent filtration applications.

8.1.8 Ultraviolet Irradiation

Ultraviolet (UV) irradiation is used to destroy bacteria and reduce organic compounds (measured as TOC) as well as destruction of chlorine and chloramines. This technique involves passing water over a UV lamp that is operating at a specific wavelength of energy.

Bacteria require a dosage of radiation equivalent to about 10,000–30,000 microwatt-seconds/square centimeter. This can be achieved by using a 254-nanometer wavelength. This wavelength alters the DNA of microbes, causing them to be unable to reproduce, leading to their death.

A significant advantage of UV over use of chemical oxidizers for microbial control is that no trihalomethane (THM) compounds are generated (see Chapter 8.2.1). Additionally, the need to store and feed a hazardous chemical oxidizer is avoided.

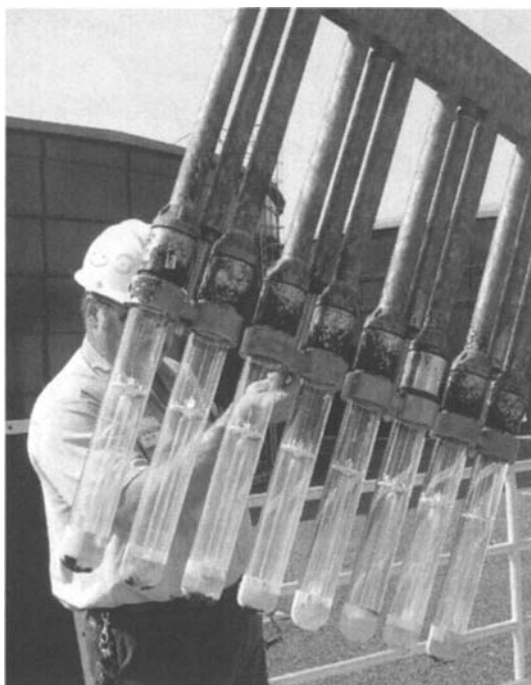


Figure 8.14 Ultraviolet light bulbs.

A limitation in using UV irradiation for microbial growth control is that there is no residual. Unlike with chemical oxidizers, there is no downstream protection against microbial growth once the treated water leaves the UV unit. Any viable material that happens to make it through the UV process is then free to grow and foul downstream equipment including the RO membranes. Also, introduction of microbes into the RO influent stream via, for example, chemical feeds, are also free to proliferate and foul piping, equipment, and membranes downstream of the introduction point.

Total organic carbon requires more energy to treat, typically around 90,000 microwatt-seconds/square centimeter. Since lower wavelengths correspond to higher energies, a wavelength of 185 nanometers is used for TOC removal. At this wavelength, organics are oxidized to form organic acids, carbon dioxide, and water.⁸

Ultraviolet radiation can also be used as an alternative to carbon or sodium metabisulfite for the destruction of chlorine and chloramines in RO feed water. The UV radiation breaks the molecular bonds of these compounds, reducing them to basic elements. For example, the typical end products of chloramine destruction using UV are chloride, hydrogen, nitrate, and ammonia. Ultraviolet radiation offers advantages over conventional dechlorination techniques, including being chemical free and without the potential for increased microbial fouling that is associated with carbon filters. Limitations of the technology include high energy requirements and capital investment.

Low-pressure mercury vapor lamps are used to produce the UV light (see Figure 8.14). An electric current is passed through an inert gas. This vaporizes mercury contained in the lamp, which then emits UV radiation. The lamp is encased in a quartz sleeve and water is in contact with the quartz. Quartz is used instead of glass because quartz does not absorb UV radiation while glass does.

Maximum contact between the water and the quartz sleeve is achieved by using plug flow with a tangential flow pattern for the water. Retention time of water in a UV unit is designed to be at least 15 minutes.

For UV to be effective, certain water conditions must be met. The water must be free of suspended solids, which can foul quartz sleeves, thereby reducing the amount of radiation reaching the water. Some UV systems include cleaning mechanisms for quartz sleeves. The water should also be free of taste, odor, iron, and manganese. Furthermore, chloride, bicarbonate, and sulfates should be reduced, as these affect the absorption of UV radiation.⁽⁹⁾ Thus, some pre-treatment is required prior to sending water to a UV system.

8.1.9 Membrane

Membrane pretreatment includes microfiltration (MF), ultrafiltration (UF), and nanofiltration (NF). Microfiltration and UF membrane processes can remove microbes and algae. However, the pores of MF and UF membranes are too large to remove the smaller, low-molecular weight organics that provide nutrients for microbes. As a result, MF and UF can remove microbes in the source water, but any microbes that are introduced downstream of these membranes will have nutrients to metabolize. Therefore, chlorination along with MF and UF is often recommended to minimize the potential for microbial fouling of RO membranes. The MF or UF membranes used should be chlorine resistant to tolerate chlorine treatment. It is suggested that chlorine be fed prior to the MF or UF membrane and then after the membrane (into the clearwell), with dechlorination just prior to the RO membranes. See Chapter 16.1 for additional discussion about MF and UF membranes for RO pretreatment.

Nanofiltration membranes are “tighter” than either MF or UF membranes but “looser” than RO membranes. They can be used to remove dissolved species, such as hardness and color. Recent developments in NF membranes have made them applicable to de-color feed water without chlorination and with minimal membrane fouling (see Chapter 16.2).

8.2 Chemical Pretreatment

Chemical pretreatment focuses on bacteria, hardness scale, and oxidizing agents. Chemicals are used to remove, destroy, inhibit, or

Table 8.7 Chemical RO pretreatment techniques and the species each technique treat.

Chemical Pretreatment	Species Addressed
Chlorine	Microbes, Total Organic Carbon, Color
Ozone	Microbes, Total Organic Carbon, Color
Antiscalants	Hardness, Silica
Sodium Metabisulfite	Oxidizers (free chlorine)
Non-Oxidizing Biocides	Microbes

Table 8.8 Oxidation-reduction potential for various compounds.¹⁰

Species	Oxidation Reduction Potential (volts)
Hydroxyl	2.8
Ozone	2.1
Peroxide	1.8
Chlorine gas	1.4
Hypochlorite	0.9

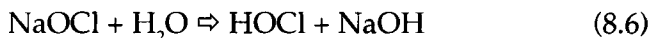
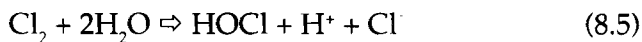
chemically reduce these species. Table 8.7 lists chemical treatments and what species they treat.

8.2.1 Chemical Oxidizers for Disinfection of Reverse Osmosis Systems

Chemical oxidizers used to disinfect RO systems include hydrogen peroxide (peroxide), halogens, and ozone. Although halogens (and specifically chlorine) are the most popular oxidizers using in conjunction with RO pretreatment, they do not have the highest oxidation-reduction potential (ORP). Table 8.8 lists the ORP for peroxide, ozone, and chlorine. As the table shows, ozone and peroxide have nearly twice the ORP or disinfection capability as chlorine. Despite the relatively low ORP, chlorine is the most commonly used disinfectant in brackish water RO pretreatment due to its ease of use and its ability to provide residual disinfection (for seawater desalination using RO, bromine (as HOBr) is predominantly used because the high bromine concentration in typical seawater would rapidly form hypobromous acid if hypochlorous acid were used).

8.2.1.1 Chlorine

Chlorine is commonly used to kill microbes in pretreatment prior to RO and to break up organics that may foul RO membranes. It is used rather than other halogens because of its higher ORP. Chlorine is available in many forms, such as chlorine gas, sodium hypochlorite (bleach), chloramines, and chlorine dioxide. Chlorine gas and sodium hypochlorite each react with water to form hypochlorous acid, as shown in Equations 8.5 and 8.6, respectively.

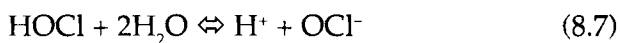


where:

NaOCl = sodium hypochlorite

HOCl = hypochlorous acid

Hypochlorous acid then dissociates in water to form hydrogen ions and hypochlorite ions (Equation 8.7):



where:

OCl⁻ = hypochlorite ion

The equilibrium in Equation 8.7 is a function of pH. Figure 8.15 shows how the presence of hydrochlorous acid, chlorine gas, and hypochlorite ion are functions of pH. Hydrochlorous acid is predominant at pH 4 – 5. At pH 7.3 or so, the concentrations of hydrochlorous acid and hypochlorite ion are the same, and above pH 8, hypochlorite ion is the only species present.

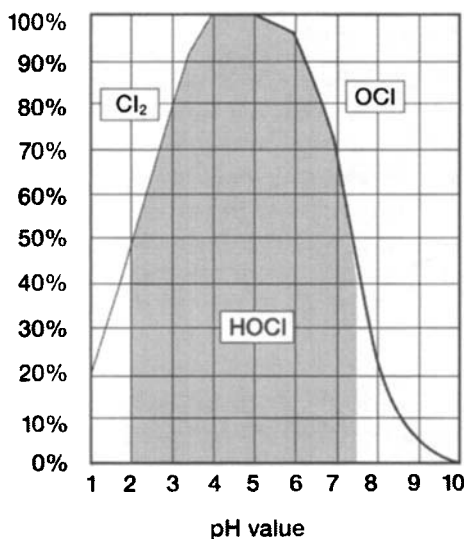


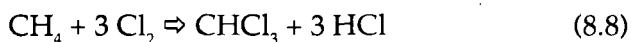
Figure 8.15 Presence of hydrochlorous acid, chlorine gas, and hypochlorite ion as functions of pH.

Hydrochlorous acid has the highest disinfecting capability of the 3 species; it is 100 times more active than hypochlorite ion.¹¹ Chlorine gas contains 100% available chlorine (hydrochlorous acid) while industrial-grade bleach has about 12% – 15% available chlorine.

The sum of chlorine gas, sodium hypochlorite, calcium hypochlorite, hydrochlorous acid, and hypochlorite ion is known as the free or free available chlorine. Most polyamide composite membranes have little tolerance for free chlorine; they can tolerate about 200 – 1,000 ppm-hrs of exposure (e.g., 200 hours at 1 ppm of free chlorine) before rejection drops to unacceptable levels. While the pretreatment to RO should have a free chlorine residual of about 0.5 to 1 ppm, the influent to the RO must be dechlorinated to bring the free chlorine concentration down to less than 0.02 ppm.

Note that some membrane manufacturers treat some of their membranes with a measured amount of chlorine prior to shipment. This brief, controlled exposure results in higher water flux with no adverse effect on rejection. These membranes are sometime referred to as “high flux” or “high capacity” membranes. Once the salt rejection decreases upon exposure to chlorine, however, the membranes are irreversibly damaged. Hence, this technique should not be attempted in the field.

Chlorination of water containing organics will create trihalomethanes (THMs):



where:

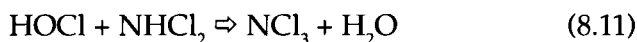
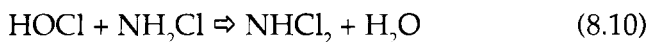
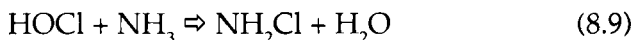
CH_4 = Methane

CHCl_3 = chloroform

Besides chloroform (the most common THM), the other three trihalomethane compounds are bromodichloromethane (CHCl_2Br), dibromochloromethane (CHClBr_2), and bromoform (CHBr_3). Trihalomethanes are considered to be carcinogenic at concentrations greater than 100 ppb. Concentration of THMs is a function of pH, temperature, contact time, and concentration of organics that react with chlorine. Concentration of THMs tends to increase with increasing pH, temperature, contact time and concentration of organics.

Control of THMs involves either eliminating the chlorine feed or the organic precursors. In cases where oxidation is required for microbial or organics control, other oxidizers should be considered.

If there is any ammonia present in the water being treated, hydrochlorous acid reacts with it to form chloramines:



where:

NH_2Cl = monochloramine

NHCl_2 = dichloramine

NCl_3 = trichloramine

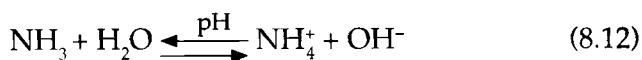
Note that trichloramine is an unstable gas that quickly dissociates into its components. The formation of the particular species of chloramines formed is dependent on pH. Trichloramine is formed at pH less than 4.4. Dichloramine is formed at pH 4.4 to 6.0. Monochloramine is the most prevalent species at pH greater than about 7.

Chloramines collectively are also known as the "combined" chlorine. The sum of the free and combined chlorine is the "total" chlorine.

In theory, the tolerance of polyamide composite membranes to chloramines is about 300,000 ppm-hrs. However, chloramines are usually in equilibrium with free chlorine, making it difficult to use chloramine in RO pretreatment, as the free chlorine will degrade polyamide composite membranes.

Although chloramines are generally not recommended by membrane manufacturers for use with polyamide composite membranes, there is some anecdotal support for the use of chloramines if the ammonia is naturally occurring in the water to be treated.⁸ In such cases, there usually is an excess of ammonia. Difficulties arise when ammonia is added to chlorine to make the chloramines. These systems tend to have more free chlorine present in equilibrium with the chloramines (see Chapter 7.11 for more discussion on this topic).

Since chloramines are created using chlorine and ammonia, there can be some ammonia present in the chloramine solution. Ammonia is a gas and as such, is not rejected by a polyamide RO membrane. Additionally, free ammonia in water will swell a polyamide RO membrane, causing it to pass more dissolved solids. The presence of the ammonia is a function of pH, as illustrated below:



pH	percent free ammonia
6.0	0
8.0	10
9.0	50

where

NH_3 = free ammonia

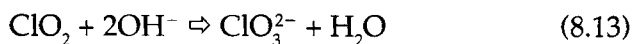
NH_4^+ = ammonium ion

It is important to keep the pH below 7.0 to assure minimal swelling of the membrane with free ammonia and minimize the impact on product quality. This pH should be maintained in the concentrate stream.

Chloramines can be removed from solution using carbon filtration, as noted in Chapter 8.1.4. However, the contact time for removal is about 4 times that of free chlorine. Chloramines can also be removed using sodium thiosulfate or bisulfite, and the reaction is fairly instantaneous. Note that with the carbon filtration removal method, some ammonia is created, which is toxic and should be considered when using an RO with chloramines for food processing and pharmaceutical applications (see Equation 8.2). However, as free chlorine is removed using sodium bisulfite, the chlorine-chloramine equilibrium can shift back to creating more free chlorine. In this case, complete removal of free chlorine cannot be assured. Carbon filters may be the best method to remove chloramines, but can take up to 30 minutes of residence time for complete reaction with the carbon. Ultraviolet radiation can also be used to destroy chloramines (see Chapter 8.1.8).

Some jurisdictions, including municipalities that treat make-up water prior to the RO pretreatment systems, have been known to switch disinfection chemicals with little or no warning. In most cases, the switch is from chlorine (or hypochlorite) to chloramines. As discussed in above, if ammonia is added to chlorine to make hypochlorite, chances are that there will be some residual free chlorine in equilibrium with the chloramines that will remain even when the chloramine is "dechlorinated." If free ammonia is present and the RO concentrate pH is greater than 7.0, RO permeate quality can be affected by the switch from free chlorine to chloramine. Any changes in effluent quality for an RO operating on municipal supply should be evaluated for the presence of chlormaine.

Chlorine dioxide is sometimes used for disinfection and organic destruction. Chlorine dioxide is a gas that does not hydrolyze into hydrochlorous acid as does chlorine. Chlorine dioxide reacts with hydroxyl under alkaline conditions to form chlorite (Equation 8.13).

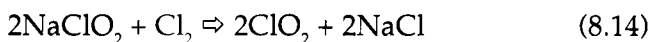


where:

ClO_2 = chlorine dioxide

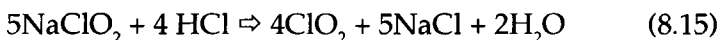
ClO_2^{2-} = chlorite

Chlorine dioxide is manufactured on site as the gas cannot be stored in compressed form; it is explosive under pressure. Several methods are used to generate chlorine dioxide. The most common method involves the reaction of chlorine gas with sodium chlorite to form chlorine dioxide and sodium chloride:



Theoretically, 1 pound of chlorine gas is required for each 2.6 pounds of sodium chlorite. In practice, however, an excess of chlorine is used to lower the pH to about 3.5 and drive the reaction to completion. In reality, the reaction never goes to completion, and there is always free chlorine in equilibrium with the chlorine dioxide. For this reason, membrane manufacturers discourage the use of chlorine dioxide for disinfecting RO systems.¹¹

Another method for generating chlorine dioxide involves the acidification of sodium chlorite with hydrochloric acid to form chlorine dioxide, sodium chloride, and water:



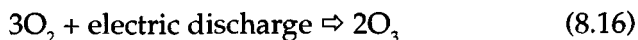
This method does not form free chlorine and therefore, could be used with RO membranes.

Chlorine dioxide forms a true solution in water; it does not hydrolyze as chlorine does. Therefore, it is very volatile (700 times more volatile than hypochlorous acid) and, therefore, can easily volatilize in RO pretreatment systems, leaving the feed water without a disinfectant residual. For this reason and the fact that most chlorine dioxide is generally in equilibrium with free chlorine, makes it unattractive for RO feed water disinfection.

8.2.1.2 Ozone

Ozone is a very powerful disinfectant. Its ORP is greater than that of chlorine. As Table 8.8 shows, the ORP for ozone is nearly twice that for hypochlorite. Ozone will also destroy a significant amount of organics as TOC.

Although ozone can be generated in a number of different fashions, the most economical method is by dielectric barrier discharge.¹⁰ This method involves the passing of a high-voltage, alternating current (6 to 20kV) through either air or pure oxygen:



When added to water, ozone quickly converts to oxygen, leaving behind no residual ozone. This makes it difficult for ozone to provide residual disinfection of RO feed water. Although no trihalomethanes are produced when using ozone, side reactions have been known to form carcinogenetic compounds such as aldehydes and phthalates.⁹

8.2.1.3 Hydrogen Peroxide

Peroxide or a combination of peroxide and peracetic acid is generally used to treat RO systems that are already contaminated with microbes. Due to its high ORP, however, a solution of only 0.2wt% peroxide is normally used (see Table 8.8). Temperature must be below 25°C and transition metals such as iron must be removed prior to treatment with peroxide to minimize oxidation of the membrane. Further, membrane should be cleaned free of deposits using an alkaline cleaner before peroxide is applied. Finally, a pH of 3 – 4 should be maintained and exposure limited to about 20 minutes for optimum result and maximum membrane life. Peroxide should not be used for storage of membrane modules.

8.2.2 Antiscalants

Sequestering agents (also known as scale inhibitors or antiscalants) are used to minimize the potential for forming scale on the surface of an RO membrane. Antiscalants work by one of three methods:

- Threshold inhibition—the ability to keep supersaturated salts in solution
- Crystal modification—the ability to change crystal shapes, resulting in soft, non-adherent scales
- Dispersion—the ability to impart a highly negative charge to the crystal thereby keeping them separated and preventing propagation.

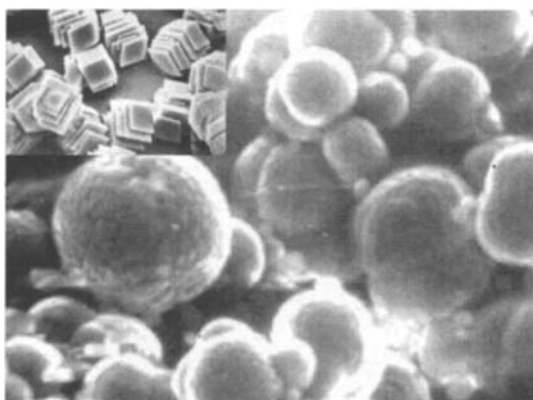


Figure 8.16 Normal crystals (inset) and crystals that have been surface-modified by an antiscalant.

Figure 8.16 shows normal crystals (inset) and crystals that have been modified by an antiscalant to inhibit their growth.

Early antiscalants used sodium hexametaphosphate (SHMP) as a threshold agent to inhibit the growth of calcium carbonate and sulfate-based scales.⁶ Most antiscalants on the market today contain sulfonate, phosphate, or carboxylic acid functional groups. Perhaps the most effective antiscalants today contain a blend of polyacrylic acid (PAA) and phosphoric acid or polyacrylate and a hydroxyethylidene diphosphonate (HEDP).¹² The polyacrylate-HEDP blends also claim to have good dispersion qualities toward silts and clays.¹² Some new inhibitors include a chelant and dispersant to keep suspended solids such as iron and manganese oxides in solution. These newer antiscalants are generally more effective than SHMP for a variety of potential scales.⁶

Antiscalants are usually fed alone for most applications. Acid feed is sometime used in conjunction with an antiscalant to control LSI for calcium carbonate scale and to control calcium phosphate scale. Antiscalants currently on the market are not generally effective at controlling calcium phosphate scale and have difficulty controlling calcium carbonate scale when the LSI is greater than about 2, depending on the manufacturer. To control calcium carbonate scale, acid is added to drop the LSI down to “acceptable” numbers. “Acceptable” numbers for LSI can range from 1.0 up to greater than 2 (recall that an LSI greater than zero is indicative of calcium carbonate scaling potential). Antiscalant is then used to provide the

balance of scale protection for calcium carbonate scale. In the case of calcium phosphate scale, enough acid should be added to decrease the calcium phosphate scaling potential to 100% or lower. A note of caution when adding acid to lower LSI and calcium carbonate/calcium phosphate scaling potential; the addition of sulfuric acid may significantly increase the potential for forming sulfate-based scales. For this reason, hydrochloric acid is preferred for pH reduction.

Antiscalant feed is typically controlled based on the feed flow rate to the RO. This type of control can lead to inconsistent dosage, as shown in Figure 8.17. Nalco Company offers 3D TRASAR® technology for RO, an alternative dosage control method (3D TRASAR, Nalco, and the logo are trademarks of Nalco Company, Naperville, IL). The 3D TRASAR® controller relies on a fluorescing molecule that allows a fluorometer to detect exactly how much antiscalant has been fed. Figure 8.18 shows how a 3D TRASAR® system controls the dosage of antiscalant to an RO.

Some antiscalants, such as those containing HEDP are sensitive to chlorine and other oxidizers.¹² They should be fed downstream of the dechlorination point. Most other antiscalants are not affected by chlorine at typical 0.5 – + 1.0 ppm concentrations.¹³

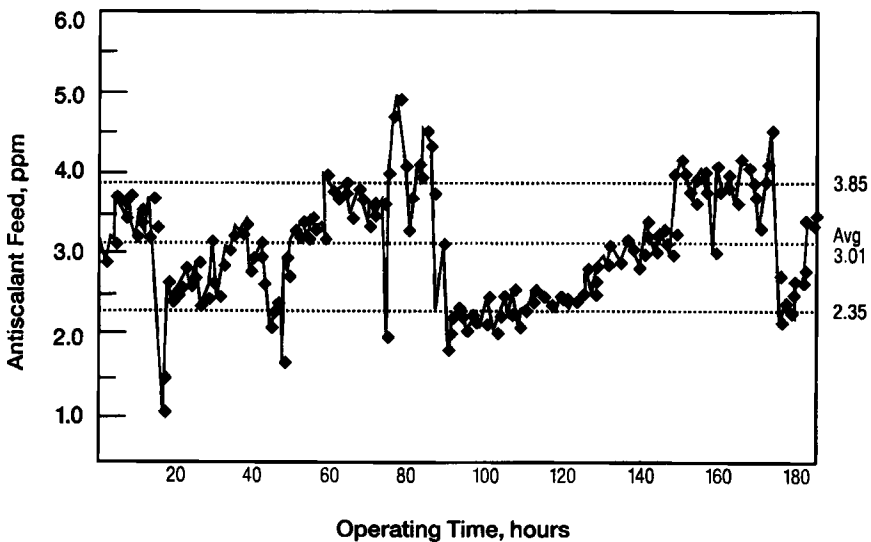


Figure 8.17 Inconsistent dosage of antiscalant based on conventional, flow-proportional control. Note the wide range of antiscalant dosages, resulting in both underfed and overfed of chemical.

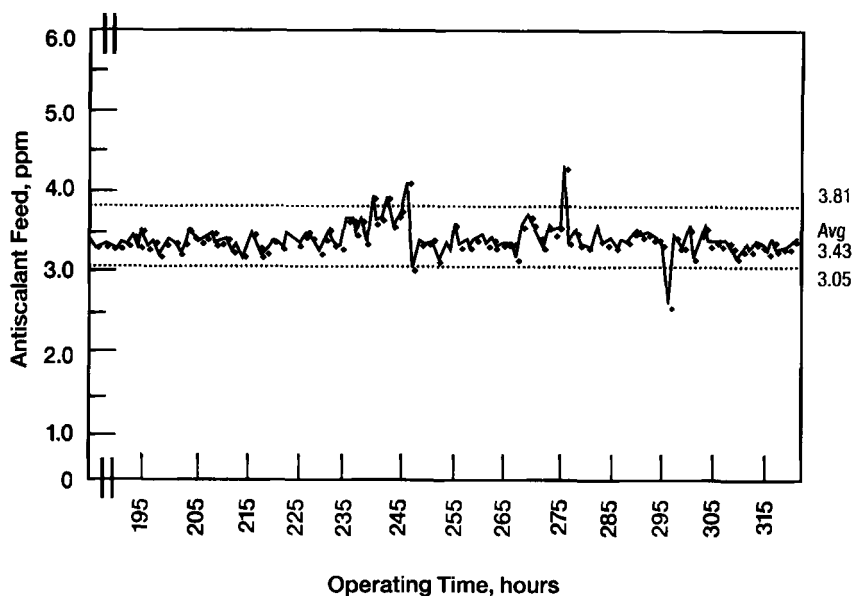


Figure 8.18 Consistent dosage control of antiscalant based on the 3D TRASAR® system. Note the relatively narrow range of antiscalant dosages resulting in few underfed or overfed episodes.

As mentioned in Chapter 8.1.1.1, overfeed of cationic coagulants can complex with negatively-charged antiscalants to co-precipitate and foul RO membranes. Care should be exercised to avoid overfeed or carryover of cationic coagulants prior to RO membranes in any case, but particularly when also feeding an antiscalant. Additionally, overdosing of antiscalants containing polyphosphate can result in calcium phosphate scale, as the polyphosphate hydrolyzes to ortho-phosphate.¹² Finally, there is a concern with microbiological contamination of antiscalant solutions. Those antiscalants containing phosphorous can accelerate the growth of microbes, as can some antiscalants in total because they themselves are food for microbes (e.g., SHMP).¹²

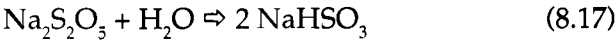
Antiscalants are completely rejected by RO membranes and, therefore, are not a concern for product quality.

8.2.3 Sodium Metabisulfite

Dechlorination of feed water to polyamide composite membranes is necessary as a polyamide membrane polymer cannot tolerate oxidizers of any kind. The options for dechlorination include activated

carbon, sodium metabisulfite chemical feed, and UV radiation. Carbon has its own set of difficulties, as described previously, and UV radiation can be capital intensive.

Sodium metabisulfite is the most commonly used technique to dechlorinate RO influent. In water, the sodium metabisulfite forms sodium bisulfite:

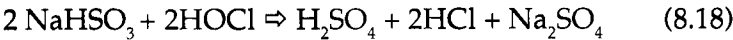


where:

$\text{Na}_2\text{S}_2\text{O}_5$ = sodium metabisulfite

NaHSO_3 = sodium bisulfite

The sodium bisulfite then reduces hypochlorous acid as follows:



Theoretically, 1.34 mg of sodium metabisulfite is required for every 1 mg of free chlorine. In practice, however, it is recommended that 2 mg of sodium metabisulfite be fed per 1 mg of free chlorine. If the sodium metabisulfite solution is 33% active, that means that about 6 mg of product should be fed per 1 mg of free chlorine.

Food-grade sodium metabisulfite that is free of impurities should be used in RO systems. The compound must not be cobalt-activated, as cobalt can catalyze the oxidation of the polyamide composite membrane in a manner similar to iron and manganese (see Chapter 7.6). Further, while the shelf life of solid sodium metabisulfite is 4 – 6 months, in solution, the shelf life depends on the concentration, as shown in Table 8.9.⁹

Sodium metabisulfite used for dechlorination should be fed down stream of all pretreatment unit operations. In other words, sodium

Table 8.9 Shelf life of sodium metabisulfite solutions of various concentrations.

Concentration of Sodium Metabisulfite (wt %)	Life
10	1 week
20	1 month
30	6 month

metabisulfite should be fed after the RO cartridge filter, if possible. This allows for protection of the cartridge filters with chlorine.

Dosage of sodium metabisulfite is typically based on Oxidation-Reduction Potentiometer (ORP). To ensure all free chlorine has been removed from RO feed water, the ORP should be controlled to read less than about 175 millivolts.

A note of caution when feeding sodium metabisulfite. If membranes are heavily fouled with heavy metals such as cobalt, iron, or manganese, residual sodium bisulfite actually converts to an oxidant in the presence of excessive oxygen. In this case, the membranes are in danger of being oxidized and destroyed.⁹

8.2.4 Non-Oxidizing Biocides

Non-oxidizing biocides are used on membranes to prevent microbial fouling. By definition, these products will not oxidize polyamide composite membranes and can be used directly on the membranes. There two most common, non-oxidizing biocides used with RO membranes: sodium bisulfite and 2,2,dibromo-3-nitrilo-propionamide or DBNPA.

8.2.4.1 Sodium Bisulfite

Sodium bisulfite can be used as a biocide on a shock feed basis. Typically, 500 to 1,000 ppm as sodium bisulfite should be fed for 30 minutes. The frequency of use should be dictated by the temperature of the water and the concentration of nutrients for microbes (warmer water and higher concentrations call for more frequent application of the bisulfite).

8.2.4.2 DBNPA

DBNPA (2,2-dibromo-3-nitrilopropionamide) can be used as a biocide on a shock feed or continuous feed basis. For shock-feed treatment, it is recommended that a concentration of about 100 ppm be fed for 30 to 60 minutes. Frequency of application depends on the degree of microbial fouling or the potential for microbial fouling, but ranges from once every 2 days to once per week. Higher temperatures, pH greater than 8.5, and the presence of residual reducing agents (such as sodium bisulfite) require higher dosages and longer contact time. Continuous treatment calls for about 1 to 2 ppm. Because some of the degradation byproducts and other ingredients in the formulations are not always rejected by the membrane, the shock treatment

is preferred, and permeate should be diverted during application as concentration of organics may increase in the system.⁹ This is particularly important in ultrapure water applications.

For potable water applications, only off-line treatment with DBNPA is recommended. This is to ensure that the single produce active concentration (SPAC) of 90 ppb of DBNPA in the permeate is not exceeded.¹⁴

Sodium bisulfite and other reducing agents can decompose the active ingredient in DBNPA formulations. Hence, it is recommended to suspend use of reducing agents during addition of DBNPA to avoid decomposition of the biocide.¹⁴

Although DBNPA is non-oxidizing, it does give an ORP response of about 400 milliVolts at 0.5 to 3.0 ppm. Chlorine gives a reading of about 700 milliVolts at 1 ppm concentration. It is recommended that the ORP set points be by-passed during addition of DBNPA.

DBNPA can also be used as a cleaner to destroy microbes within the membrane modules. Dosage is about 100 ppm for 30 minutes at pH 6.0 to 7.5. This cleaning may be followed by an alkaline cleaning to help remove any biofilm that may be present (see Chapter 13.2 for more details on membrane cleaning formulations).

Stainless steel injection quills should not be used with DBNPA as they may corrode.

8.2.4.3 *Other Non-Oxidizing Biocides*

Isothiazolone has also been used as a non-oxidizing biocide for RO applications. However, the residence time required is much longer than for DBNPA. For example, a dosage of 50 to 100 ppm requires a 4-hour contact period. Thus, isothiazolone is not recommended for shock feed or continuous feed, but is recommended for cleaning events. Isothiazolone is more effective than DBNPA in high-organic waters.

Quaternary germicides, phenolic compounds and iodine are not recommended as sanitizing against for polyamide membranes as these compounds can cause losses in water flux through the membrane.¹⁵

8.3 **Combination Mechanical Plus Chemical Pretreatment—Lime Softening**

Lime softening is used to remove the following species from water:

- Calcium
- Iron

Table 8.10 Effluent from cold, warm, and hot lime softeners.

Species	Raw Water	Cold Lime	Cold Lime-Soda	Warm Lime	Hot Lime	Hot Lime-Soda
Total Hardness (ppm CaCO_3)	250	145	81	70	120	20
Calcium (ppm CaCO_3)	150	85	35	30	115	15
Magnesium (ppm CaCO_3)	100	60	46	40	5	5
P Alkalinity (ppm CaCO_3)	0	27	37	24	18	23
M Alkalinity (ppm CaCO_3)	150	44	55	40	28	40
Silica (ppm)	20	19	18	15	1 – 2*	1 – 2*
pH	7.5	10.3	10.6	10.4	10.4	10.5

* Silica removal to this concentration may require the additional feed of magnesium oxide with sludge recirculation.

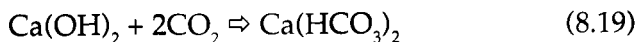
- Turbidity
- Organics
- Silica
- Magnesium
- Manganese
- Color
- Oil

Lime softening can be conducted cold (ambient), at warmer temperatures, or hot, where steam is used to heat the process. The differences among the three options are in the removal of hardness, alkalinity, and silica. Table 8.10 lists approximate effluent from cold, warm, and hot lime softeners.¹⁶

8.3.1 Cold Lime Softening

Cold lime softening is conducted at ambient temperatures and involves feeding calcium as lime (Ca(OH)_2) to precipitate out calcium

carbonate:



where:

Ca(OH)_2 = lime

$\text{Ca(HCO}_3)_2$ = calcium bicarbonate

CaCO_3 = calcium carbonate precipitate

Calcium can be reduced to about 35 – 50 ppm in this manner.

The removal of other species requires the addition of sodium aluminate ($\text{Na}_2\text{Al}_2\text{O}_4$). Reduction in the concentration magnesium is only about 10%. The addition of sodium aluminate also helps with the reduction in silica, as the magnesium precipitate, magnesium hydroxide, adsorbs silica. More complete removal requires the addition of soda ash (Na_2CO_3) and warmer temperatures, as shown in Table 8.10.

Cold lime and lime-soda softening is conducted in a solids-contact clarifier (see Chapter 8.1.1.). Cold lime softeners are typically followed by filtration and sodium softening to remove the balance of hardness from the water.

The cold lime process is slow, and gets slower as the temperature gets colder. For very cold waters, there is the danger of post-precipitation elsewhere in the facility, as some of the reactions will not be completed in the lime softener, and will continue into the transfer lines.

8.3.2 Warm Lime Softening

Warm lime softening is conducted at 120 – 140°F, where the solubilities of calcium and magnesium are reduced. Temperature control is critical. A change of as little as 4°F can cause carryover of softener precipitates. Conventional cold lime softening equipment is used for warm lime softening.

8.3.3 Hot Process Softening

Hot lime softening (also known as hot process softening) is conducted at 227 – 240°F. At these temperatures, the lime softening

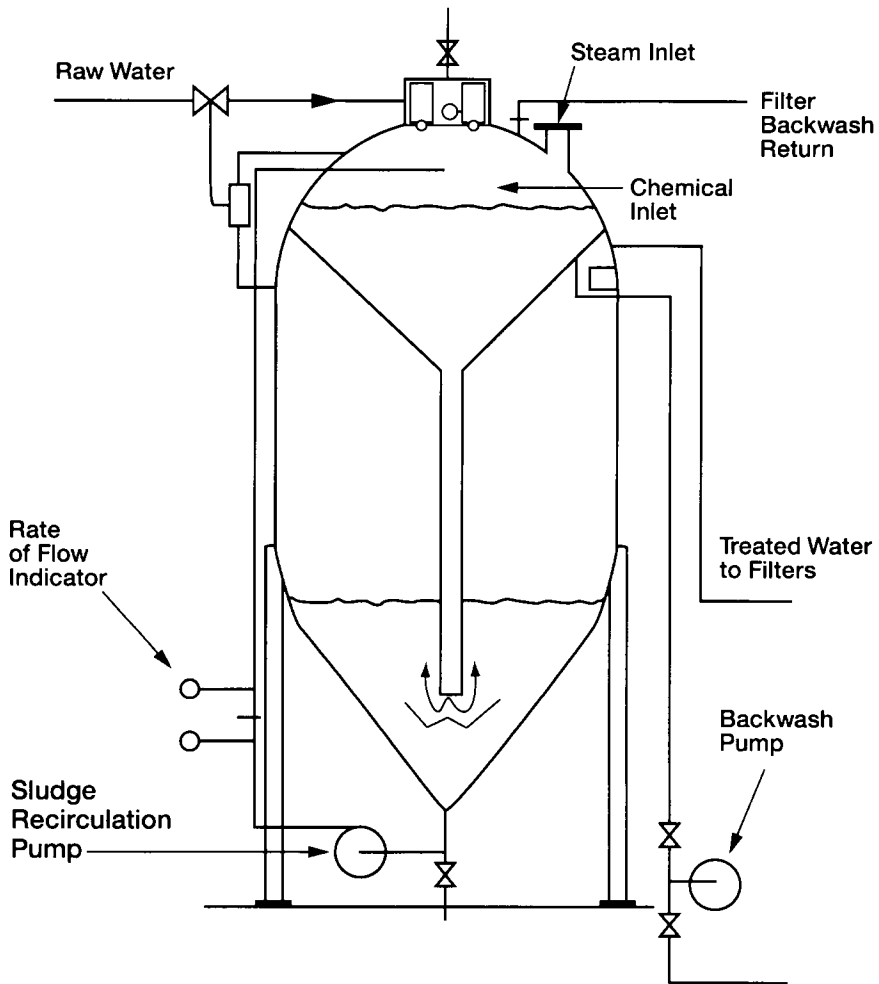


Figure 8.19 Sludge-blanket (upflow) hot process softener. *Courtesy of Res-Con, Inc.*

reactions go to completion. Calcium can be reduced to about 8 ppm while magnesium can be reduced to about 2 – 5 ppm. Furthermore, silica can be reduced to 1 – 2 ppm.

Steam is used to heat the process. Figure 8.19 shows the cross section of a sludge-blanket (upflow) hot process softener. The operations of the sludge blanket hot process unit is similar to that for sludge-blanket clarifiers (see Chapter 8.1.1.1). Figure 8.20 shows the cross section of a downflow hot process softener. The downflow units rely on recirculation pumps to provide sludge contact.

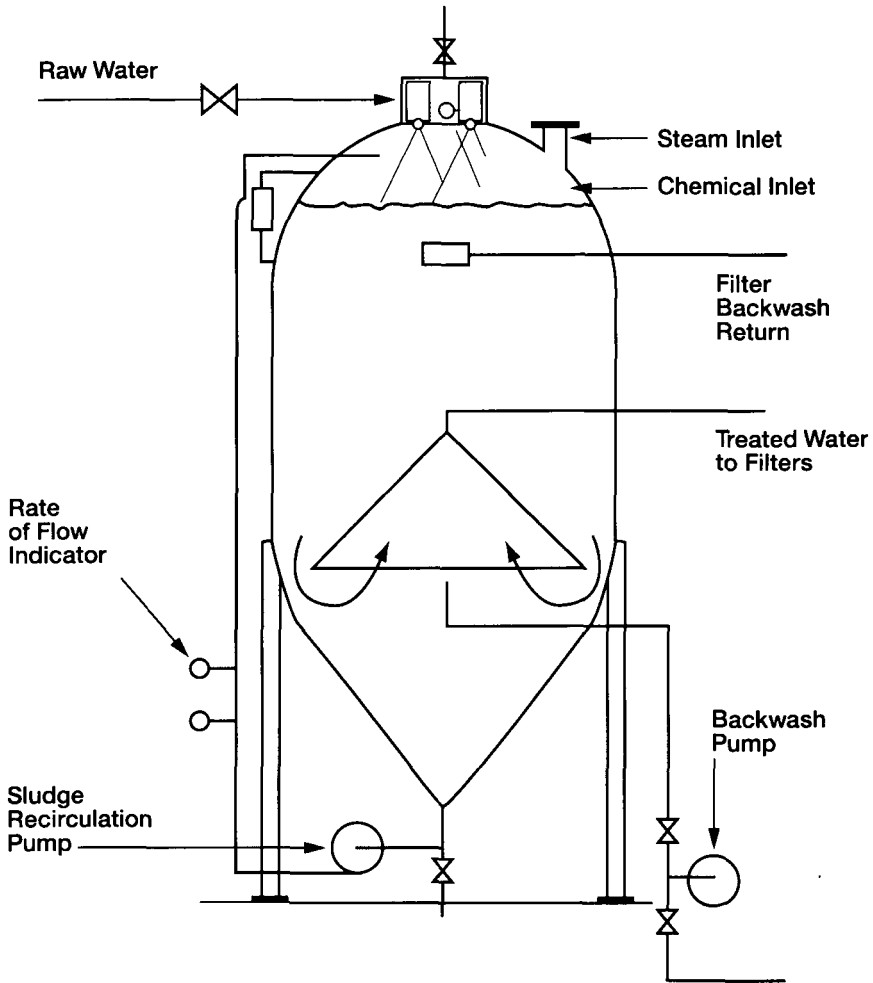


Figure 8.20 Downflow hot process softener. Courtesy of Res-Con, Inc.

Silica reduction is accomplished by adsorbing silica on the magnesium hydroxide precipitate. If not enough magnesium is present, magnesium oxide (MgO) can be added to provide the necessary adsorption sites.

8.4 Sequencing of Pretreatment Technologies

Although sequencing of pretreatment technologies is always site specific, there are some generalizations that can be made. Figure 8.21

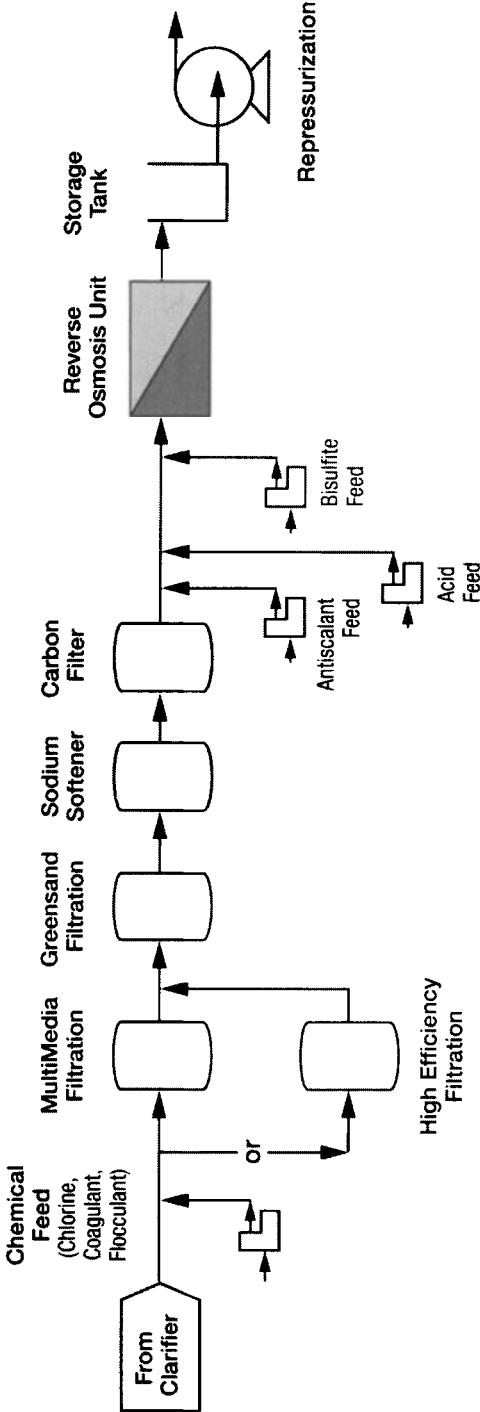


Figure 8.21 Sample reverse osmosis and pretreatment process flow diagram.

shows a typical process flow diagram that includes many of the pre-treatment technologies described above.

- **Chemical feed:** this includes chlorine, coagulants, and flocculants. The chemical feeds disinfect the water and prepare it for solids and/or hardness removal.
- **Media filtration:** this step includes multimedia filtration or high-efficiency filtration, and reduces the concentrations of suspended solids, turbidity, and SDI.
- **Greensand filtration:** This step removes soluble iron and manganese (and sometime hydrogen sulfide). This type of filtration is typically only used on well water sources. In some cases, media filtration is not required prior to greensand filtration.
- **Sodium softener:** The softener removes hardness and any residual soluble iron from the RO feed water.
- **Carbon filtration:** carbon removes chlorine and organics from RO feed water.
- **Bisulfite feed:** bisulfite is added to eliminate free chlorine, if carbon is not used.
- **Acid feed:** acid is added to reduce the LSI and calcium carbonate scaling potential, if required.
- **Antiscalant feed:** antiscalant is used to minimize the potential for scaling the membranes.
- **Reverse osmosis:** the RO unit removes the bulk of the dissolved solids from the feed water. It is typically followed by a storage tank and repressurization for transport on to post-treatment or process.

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SYSTEM DESIGN

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Design Considerations

Operating conditions affect the performance of an RO system. These conditions include:

- Feed water quality and source
- Temperature
- Pressure
- Feed water flow
- Concentrate flow
- Beta
- Recovery
- pH
- Flux

The effects of these conditions on the performance of an RO system are discussed in this chapter.

9.1 Feed Water Quality

Feed water quality and its tendency to foul has a significant impact on the design of an RO system. Selection of the design flux, feed water and reject flows (and hence, the array), and salt rejection is influenced by the feed water quality.

9.1.1 Feed Water Source

The feed water source has a great impact on the potential of the water to foul an RO membrane. High-quality source water, such as well water with SDI less than 3, has a lower chance of fouling an RO membrane than a lower-quality source water, such as surface water with an SDI of 5. An RO system designed to operate on higher-quality source water can be designed with a higher flux than one operating on a lower-quality source water. This is because a higher the flux rate brings contaminants (suspended solids, hardness) to

the surface of the membrane faster than would a lower flow rate. These contaminants then collect in the concentration polarization boundary layer at the membrane surface, which leads to accelerated fouling or scaling of the membrane. Hence, the higher the concentration of suspended solids and hardness in the feed water, the lower the flux should be to reduce the potential for fouling and scaling the membranes.

Table 9.1 lists the recommended average and conservative flux rates for various feed water source qualities.^{1,2} As the table shows, an RO system operating on well water could be designed with a flux as high as 14–16 gfd, while a surface water with SDI less than 5 should only have a flux of 10–12 gfd. In other words, for a given product flow rate, a well water-based RO system can have a 14% smaller RO system than the surface water RO due to the higher allowable flux. This not only reduces capital and operating costs, but also results in additional operating cost savings due to reduced membrane cleaning and replacement frequencies because of the higher quality feed water.

Table 9.1 Recommended flux rates as a function of feed water source, as adapted from Dow Water and Process Solutions and Hydranautics.^{1,2}

Feed Water Source	Silt Density Index	Average Flux, gfd*	Conservative Flux, gfd*
RO Permeate	< 1	21 – 25	22
Well Water	< 3	14 – 16	14
Surface Supply	< 3	12 – 14	12
Surface Supply	< 5	10 – 12	10
Secondary Municipal Effluent— Microfiltration Pretreatment**	< 3	10 – 14	10
Secondary Municipal Effluent— Conventional Pretreatment	< 5	7 – 12	7

* For 8-inch diameter, brackish water membrane modules

** Microfiltration pore size < 0.5 microns.

Feed water source also influences the design array of the RO unit. This is because the feed water flow and concentrate flow rates are also determined based on feed water quality. Higher feed water quality allows for higher feed flows and lower concentrate flows to be employed. Higher feed water flows and lower concentrate flows reduce the number of membrane modules required in the RO system.

Tables 9.2 and 9.3 list the recommended feed water and concentrate flow rates, respectively, as functions of feed water source quality.¹ Higher feed water flow rates result in water and its contaminants being sent to the membrane more rapidly, leading to faster rates of fouling and scaling. As Table 9.2 shows, an RO operating on a well water source can have a feed flow rate as high as 65 to 75 gpm per pressure vessel, while a surface water source RO should not exceed 58 to 67 gpm per pressure vessel. The well water RO would require 12% fewer pressure vessels than the surface water RO.

Table 9.2 Recommended feed water flow rate as a function of feed water source for brackish water membranes, as adapted from Dow Water and Process Solutions.¹

Feed Water Source	Maximum Feed Flow Rate for 365 ft² Modules gpm	Maximum Feed Flow Rate for 400 and 440 ft² Modules, gpm
RO Permeate	65	75
Well Water	65	75
Surface Supply	63	73
Surface Supply	58	67
Secondary Municipal Effluent—Microfiltration Pretreatment*	52	61
Secondary Municipal Effluent—Conventional Pretreatment	52	61

* Microfiltration pore size < 0.5 microns

Table 9.3 Recommended concentrate flow rates as a function of feed water source for brackish water membranes, as adapted from Dow Water and Process Solutions.¹

Feed Water Source	Minimum Concentrate Flow Rate for 365 ft ² Modules, gpm	Minimum Concentrate Flow Rate for 400 and 440 ft ² Modules, gpm
RO Permeate	10	10
Well Water	13	13
Surface Supply	13	13
Surface Supply	15	15
Secondary Municipal Effluent—Microfiltration Pretreatment*	16	18
Secondary Municipal Effluent—Conventional Pretreatment	18	20

* Microfiltration pore size < 0.5 microns

The opposite is true for the concentrate flow rate. Here, the slower the flow rate, the thicker the concentration polarization boundary layer, and the greater the chance for fouling or scaling the membranes. Table 9.3 shows that an RO operating on relatively clean well water should have a concentrate flow rate of not less than 13 gpm per pressure vessel, while an RO operating on high-solids surface water should have a concentrate flow rate of not less than 15 gpm per pressure vessel.

9.1.2 Total Dissolved Solids

The total dissolved solids (TDS) concentration affects both the system flux and the salt rejection of an RO system. Figures 9.1 and 9.2 shows the effect of TDS on flux and rejection, respectively, under conditions of constant pressure.¹ As feed TDS increases, the driving force for water decreases (under constant applied pressure), due to the increase in osmotic pressure of the feed. This results in

a decrease in system flux. As the driving force for water decreases, the amount of water passing through the membrane relative to the amount of salt decreases, resulting in a higher TDS concentration in the permeate. Thus as shown in Figure 9.2, the rejection is lower (and salt passage is higher) at higher feed water TDS.

9.1.3 Calcium and Natural Organic Matter

Divalent cations, particularly calcium, have been shown to enhance fouling of membranes with natural organic matter (NOM).³ Because of its acidic nature, NOM can form complexes with dissolved metal ions. The strongest bonds occur with calcium. This is a function of the size of the metal ion, its electronic charge, and the

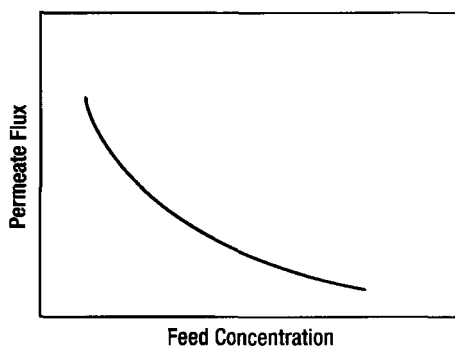


Figure 9.1 Reverse osmosis membrane flux as a function of feed water total dissolved solids. Assumes constant applied feed pressure.

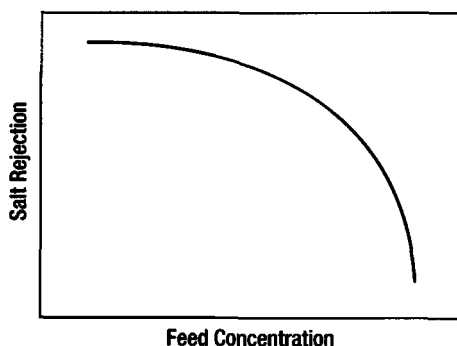


Figure 9.2 Reverse osmosis membrane rejection as a function of feed water total dissolved solids. Assumes constant applied feed pressure.

energy it takes to break the shell of water molecules that typically hydrate metal ions in water. When the hydration shell breaks away from calcium, several negatively-charged NOM molecules can simultaneously attached to the same calcium ion, creating a much larger particle. These particles are responsible for forming the bio-fouling layers on membrane surfaces.⁴

Factors that affect fouling with NOM-calcium complexes include permeate flux and cross flow rate (see Chapters 3.3 and 9.4). At higher flux though the membrane, the concentration of calcium increases in the concentration polarization boundary layer at the membrane surface, as described above. Lower cross flow rates also increase the concentration of calcium in the boundary layer. The increases concentration of calcium at the membrane surface enhances the fouling of the membranes by the NOM-calcium aggregates.⁵

9.1.4 Chemical Damage

Chemical damage occurs when a contaminant in the feed water is incompatible with the polymer comprising the membrane, the microporous support, or the fabric support. Besides oxidizers that degrade the crosslinking of a thin-film membrane, there are a variety of chemicals that swell or dissolve the polysulfone microporous support, including the following compounds.

- Ketones
- Aldehydes
- Esters
- Strong ethers
- Aromatic hydrocarbons such as benzene, toluene, xylene, and diesel fuel and gasoline.
- Solvents such as dimethyl formamide (DMF), dimethyl sulfoxide (DMSO), dimethyl acedimide (DMAc) to name a few.

Low-molecular weight solvents such as methanol, propanol, and isopropanol are considered acceptable.

9.2 Temperature

Temperature influences system flux and rejection performance. Figures 9.3 and 9.4 shows the effect on water flux and salt

rejection, respectively, under conditions of constant pressure and at temperatures less than 45°C. As shown in Figure 9.3, water flux is linearly proportional to the water temperature. For every 1°C change in temperature, there is a 3% change in water flux. This occurs because the lower viscosity of warmer water allows the water to flow more readily through the membranes.⁶ On the other hand, salt rejection decreases slightly with increasing temperature. Salt diffusion through the membrane is higher at higher water temperature (the salt transport coefficient shown in Equation 4.2 is a function of temperature).

In practice, temperature changes are dealt with by adjusting the operating pressure: lower pressure in the warmer summer months and higher pressure in the colder winter months. If there are significant variations in temperature between summer and winter, a

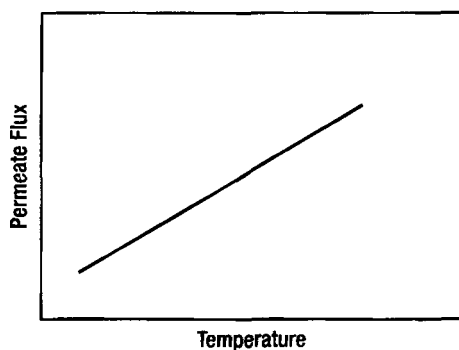


Figure 9.3 Reverse osmosis membrane water flux as a function of temperature. Assumes constant applied feed pressure and less than 45°C temperature.

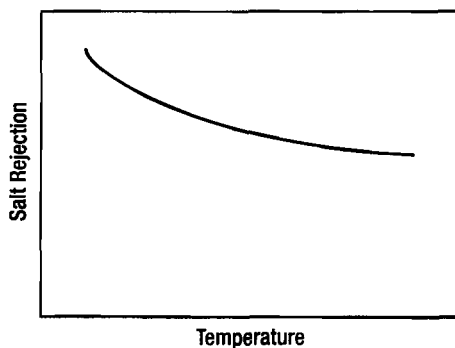


Figure 9.4 Reverse osmosis membrane salt rejection as a function of temperature. Assumes constant applied feed pressure and less than 45°C temperature.

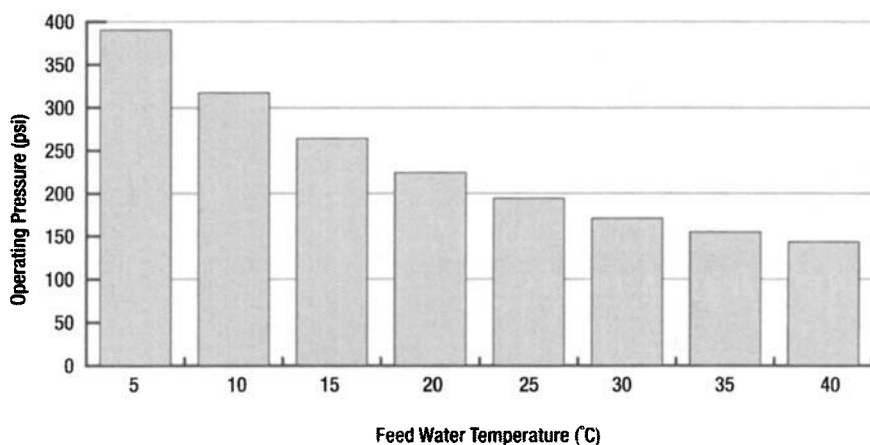


Figure 9.5 Operating pressure as a function of feed water temperature. Assumes 2000 ppm TDS at pH 7.6 feed water, 75% recovery, 3:2:1 array with 3 membranes per pressure vessel, FilmTec BW30-400/34 membranes.

variable frequency drive (VFD) can be used to adjust the speed of the feed pump motor to run according to the water temperature (see Chapter 6.2). A VFD can save considerable energy in the summer months. Figure 9.5 shows how the operating pressure changes with changing feed water temperature for a brackish water system.

At temperatures greater than 45°C, the structure of the membrane itself changes. The membrane anneals, meaning it gets denser. As a result, it becomes more difficult to force water through the membrane.⁷ At temperatures greater than 45°C, flux goes down and rejection goes up, assuming constant driving pressure.

9.3 Pressure

Operating pressure directly affects water flux and indirectly affects salt rejection. Figures 9.6 and 9.7 show the effect of pressure on flux and rejection, respectively. Because operating pressure directly affects the driving force for water across the membrane, higher pressure will result in higher flux (see Equation 4.1). Salt transport, however, is unaffected by pressure (see Equation 4.2). So, the same amount of salt passes through the membrane at low or at high feed water pressure. However, because more water has passed through the membrane at higher pressure, the absolute salt concentration in the permeate is

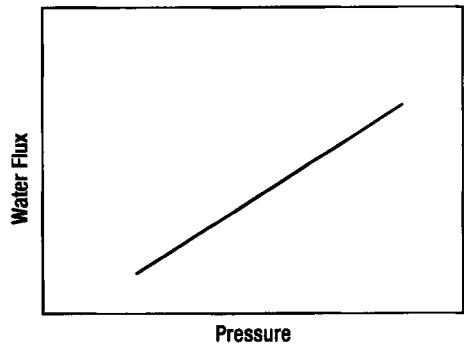


Figure 9.6 Reverse osmosis membrane water flux as a function of pressure.

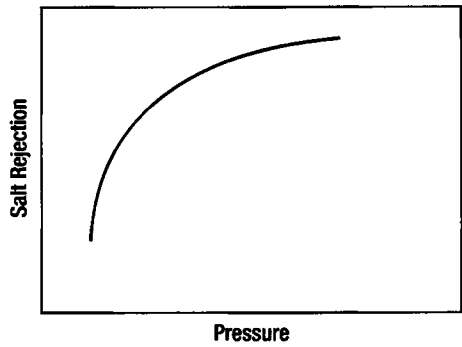


Figure 9.7 Reverse osmosis membrane salt rejection as a function of pressure.

lower, so it appears as if the salt passage decreases and the salt rejection increases as pressure increases, as shown in Figure 9.7.

9.4 Feed Water Flow

The feed water flow through an RO system should be dictated by the water source, as described in Chapter 9.1. The “cleaner” the water source, the higher the feed water flow may be, resulting in smaller systems and lower overall cost of operation.

Table 9.2 listed the recommended feed flow rates as a function of water source.¹ At higher feed water flow rates, contaminants such as colloids and bacteria that may be present in the source water, are sent to the membrane more rapidly, resulting in faster fouling of the membrane. This is why lower flow rates are recommended for water sources that contain high concentrations of contaminants.

9.5 Concentrate Flow

The recommended concentrate flow rate is a function of feed water source, as described in Chapter 9.1. The “cleaner” the water source, the lower the concentrate flow may be, resulting in smaller systems and lower overall cost of operation.

Table 9.3 listed the recommended concentrate flow rates as a function of water source.¹ At lower concentrate flow rates, good cross-flow velocity is not maintained, and contaminants, such as colloids and scale-formers, have a much greater chance of fouling or scaling a membrane. This is because the concentration polarization boundary layer is thicker at lower cross-flow velocities than it would be at higher concentrate flow rates. Since the bulk concentration of contaminants toward the concentrate end of the pressure vessel can be 3, 4, or even 5 times the concentration as found in the feed water, and since the concentration of contamination is even higher in the boundary layer, the potential for fouling or scaling a membrane can be very high at low concentrate flow rates.

9.6 Beta

Beta is the ratio of the concentration of a species at the membrane surface to that in the bulk solution, as described in Chapter 3.5. Beta is not a property of the membrane, nor does the designer of the RO system directly select it. It is a function of how quickly the influent stream is dewatered through the RO system. Hence, Beta is a consequence of the system design that is selected.

Beta affects both the flux through an RO membrane and the salt rejection. The increase in Beta due to concentration polarization at the membrane surface results in increased osmotic pressure and decrease in water flux, as shown in Equation 9.1 (modified Equation 4.1). Salt passage also increases, as shown in Equation 9.2 (modified Equation 4.2).

$$J_w = A (\Delta P - \beta \Delta \pi) \quad (9.1)$$

$$J_s = K (\beta C_{A2} - C_{A3}) \quad (9.2)$$

where:

J_w = water flux

A = Water permeability coefficient

ΔP = applied pressure driving force

β = Beta

$\Delta\pi$ = osmotic pressure of feed – concentrate solution

J_s = salt flux

K = salt permeability coefficient

C_{A2} = molar concentration of solute in boundary layer

C_{A3} = molar concentration of solute in permeate

Figures 9.8 and 9.9 shows how Beta affects flux and salt passage (rejection), respectively, for two different brackish water concentrations (assumes membrane will deliver 20 gfd at 400 psi with a rejection of 99% at Beta equal to one (no concentration polarization)).⁸ From the Figures, it is shown that at Beta values greater than about 1.1, the water flux and salt passage (rejection) are significantly affected by Beta. Also shown is that the effect of Beta on performance is more pronounced at higher TDS feed water than with lower TDS feed water.

In reality, Beta for RO systems is greater than 1.0, and hence, concentration polarization always exists. While concentration polarization cannot be eliminated, it can be minimized by judicious RO system design:

- Beta can be changed by adjusting the permeate back-pressure on each individual stage. This can be accomplished by adding flow restrictors into the permeate

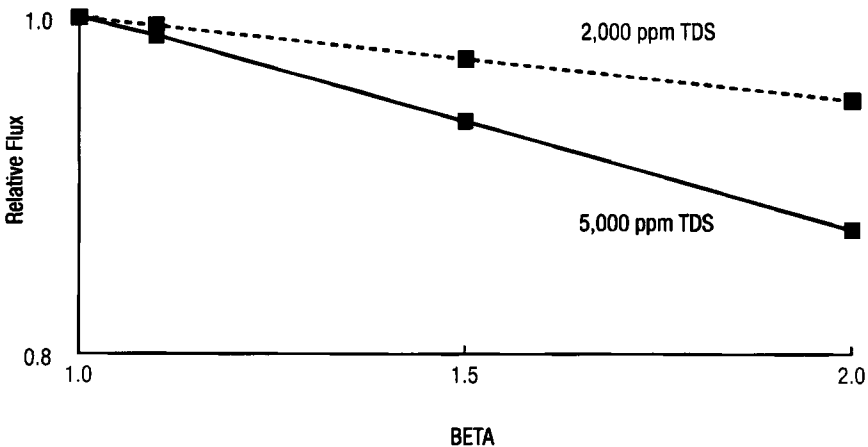


Figure 9.8 Beta's effect on membrane water flux for two different brackish water concentrations.

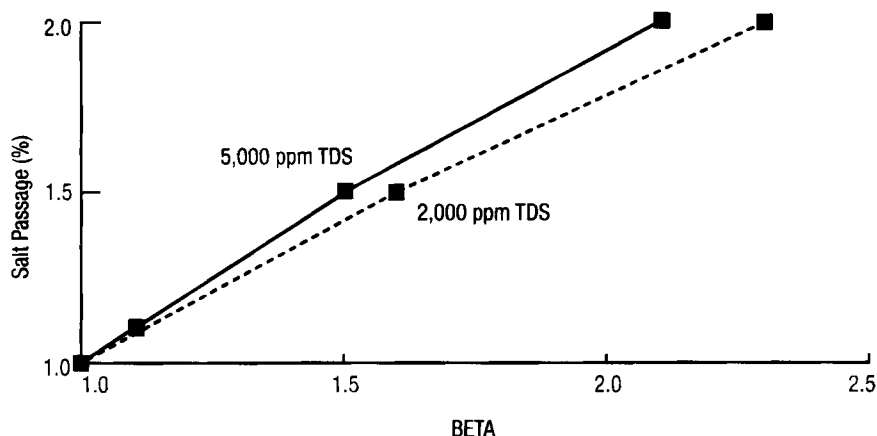


Figure 9.9 Beta's effect on salt passage for two different brackish water concentrations.

pipeline. Increasing the backpressure decreases water flux (the rate at which water is removed from the influent stream) which, in turn, decreases Beta.

- Adjusting the array of the RO system can also change Beta. Increasing the number of pressure vessels in a subsequent stage will decrease the applied pressure required by the prior stage, thereby decreasing the flux and Beta for that prior stage.
- Adherence to recommended concentrate flow rates and membrane module recovery can also minimize Beta and the effects of concentration polarization.

Conventional wisdom calls for Beta values less than 1.2 in an RO design to minimize membrane fouling and scaling.⁹ Table 9.4

Table 9.4 Hydranautics' recommendation for beta values as a function of feed water quality.²

Beta (Individual Module)	RO Permeate	Brackish Well Water	Brackish Surface Water	Tertiary Wastewater
Conservative	1.30	1.18	1.18	1.18
Maximum	1.70*	1.2	1.2	1.2

* aggressive design, 1.40 is typical.

lists Hydranautics' recommended beta values as a function of feed water quality.²

9.7 Recovery

Reverse Osmosis recovery affects the overall water flux and salt rejection as shown in Figures 9.10 and 9.11 respectively. As the recovery increases, the water flux decreases slowly until the recovery is so high that the osmotic pressure of the feed water is as high as the applied pressure, in which case, the driving force for water through the membrane is lost and the flux ceases.

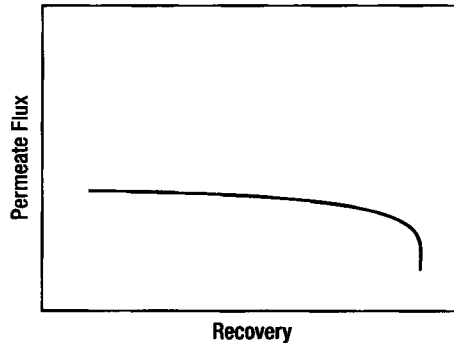


Figure 9.10 Reverse osmosis membrane water flux as a function of recovery. Assumes constant applied feed pressure.

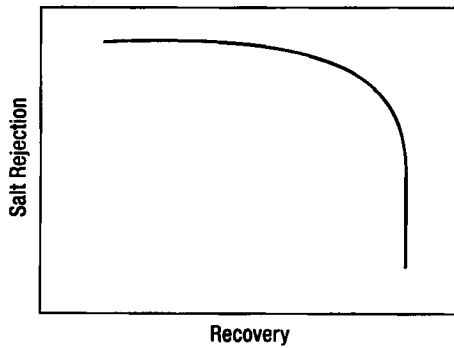


Figure 9.11 Reverse osmosis membrane salt rejection as a function of recovery. Assumes constant applied feed pressure.

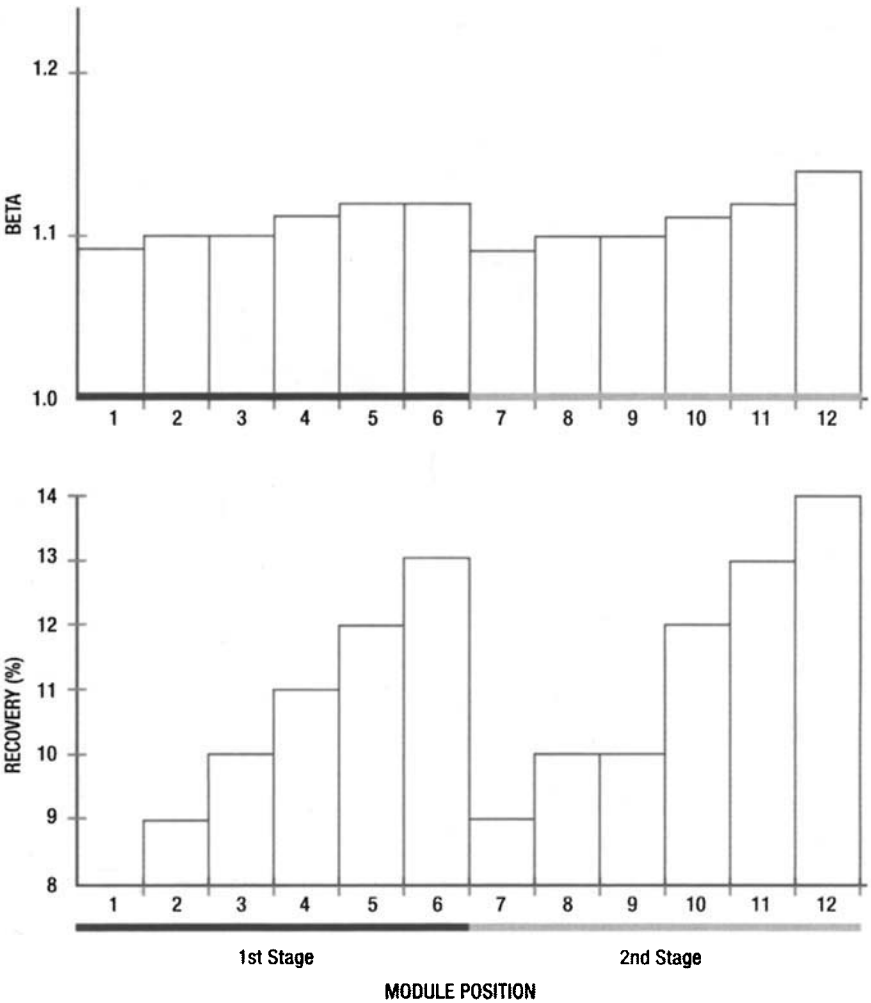


Figure 9.12 Individual membrane module recovery and rejection as a function of position in a 2-stage RO system with 6 modules per pressure vessel.

The drop in water flux affects the apparent salt rejection. As the osmotic pressure of the feed/concentrate stream approaches the applied pressure, the driving force for water is decreased, but the driving force for salt is unaffected (see Equation 4.2, which shows that the solute flux is not a function of pressure driving force). Hence, less water passes through the membrane relative to the amount of salt passing through the membrane. Thus, it appears as if the salt passage increases and salt rejection decreases with increasing recovery. Salt rejection becomes 0% at about the same time that the flux ceases.

Recovery through *individual* membrane modules changes, based on the position of the module in the pressure vessel. Most spiral wound membrane modules operate with individual module recoveries ranging from 10% to 15%, with an average of 11% to achieve 50% recovery in a single, 6-module pressure vessel stage. The module at the feed end of the pressure vessel typically exhibits the lowest recovery of all modules in the pressure vessel. The module at the concentrate end of the pressure vessel operates at the highest recovery in the vessel. This is because as more water is recovered through the pressure vessel, the percent of the total feed to a given module recovered by that module gets larger. Figure 9.12 shows this effect. Also shown in the figure is how Beta for individual modules changes through the pressure vessel. High recovery conditions leads to increased rates of fouling and/or scaling due to a higher concentration of species in the boundary layer, hence the higher Beta.

The designer of an RO system has at their discretion, the ability to manipulate the recovery of the system to minimize the potential for scaling of the membrane, both on the system and individual module level. As system recovery increases, the concentration of salts in the feed/concentrate stream increases to the point of saturation for some species near the outlet end of the pressure vessel. If acid and antiscalants are not effective (or cannot be used), recovery is typically decreased to minimize the scaling potential. Higher recovery can also result in too much water being removed in the lead modules in the pressure vessel, making them more prone to fouling and scaling. Adjusting the recovery down will increase both the flux and rejection as described in Chapter 9.6. Adjusting the recovery so that it is higher (to minimize concentrate waste) is only recommended when the feed water is relatively free of suspended solids and scale formers.

9.8 pH

pH affects the stability of both polyamide composite and cellulose acetate membranes. Cellulose acetate membranes are stable over a pH range of 4 to 6, due to hydrolysis at higher and lower pH. (Hydrolysis is a chemical reaction where a compound is broken down by reaction with water.) Polyamide composite membranes also react with water, but the pH range of nominal application is

much broader, ranging from as low as 2 to as high 11, depending on the specific membrane and manufacturer. Acceptable operating pH is a function of temperature, with higher temperatures requiring narrower pH ranges of operation.

pH also affects the rejection capabilities of polyamide composite membranes. Rejection of most species is highest at about pH 7.0 – 7.5, as shown in Figure 9.13.¹⁰ Rejection drops off at higher and at lower pH, but the drop off is very gradual at lower pH. The reason for this phenomenon is not clearly defined in the literature, but most likely stems from the ionic state of the ions being rejected, as well as some changes on the molecular level with the membrane itself. The flux through a polyamide composite RO membrane is relatively constant over the range of pH, as shown in Figure 9.14.¹⁰

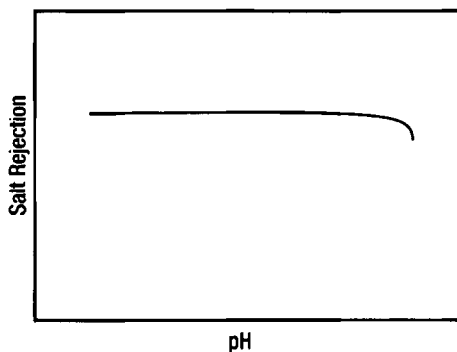


Figure 9.13 Reverse osmosis membrane salt rejection as a function of pH. Assumes constant applied feed pressure.

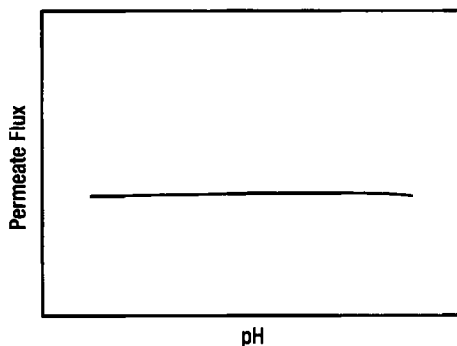
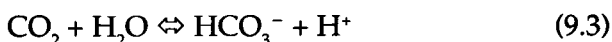


Figure 9.14 Reverse osmosis membrane water flux as a function of pH. Assumes constant applied feed pressure.

Because of the carbon dioxide present in most waters, the pH of RO product water is generally lower than the pH of feed water, unless the carbon dioxide is completely removed from the feed water. If carbon dioxide is present in feed water, it will be present in permeate, as gases are not rejected by RO membranes (see Chapter 3.2). However, the membrane rejects carbonate and bicarbonate. Passage of carbon dioxide upsets the equilibrium among these compounds in the permeate. A new equilibrium occurs in the permeate, hence lowering its pH:



9.9 Flux

Flux determines the overall size of the RO system in terms of membrane area required to achieve the desired separation. As discussed in Chapter 9.1.1, the water flux for a given application should be based on the feed water source. “Cleaner” source water allows for higher flux, which, in turn, means less membrane area is required to achieve the desired separation.

Water flux is affected by several operating variables, as discussed in this chapter. In summary:

- Flux is directly proportional to operating pressure.
- Flux is directly proportional to water temperature.
- Flux decreases slightly as recovery increase until the osmotic pressure of the feed water equals the driving pressure, at which point productivity ceases.
- Flux decreases with increasing feed concentration of dissolved solids.
- Flux is relatively constant over a range of pH, although for some newer polyamide membranes, flux is also a function of pH.¹¹

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RO Design and Design Software

Sound RO system design incorporates all issues discussed in Chapters 1 – 9. Perhaps the most important considerations when developing a new design are the following:

- Water flux (see Chapter 3.4)
- Feed flow rate per pressure vessel (see Chapter 9.4)
- Concentrate flow rate per pressure vessel (see Chapter 9.5)
- Beta (see Chapter 9.6)
- Scaling indexes (see Chapter 3.8)

Adherence to the recommended guidelines for these variables can determine the success of a design. This chapter covers the basics of designing an RO system, including the use of design software available from various membrane manufacturers.

The starting point in any RO design is the water flux. The desired water flux should be selected by the designer based on the feed water source and quality (see Chapter 3.4 and Table 3.3).

Once the flux has been selected, the other variable that the designer needs to determine is which membrane to use and how much membrane area per module is suitable for the particular application. As discussed in Chapter 4.3.3, 8-inch diameter spiral wound modules are available with various membrane areas. For example, the Dow Water and Process Solutions FilmTec FT-30 membrane is available with 365 ft² or 400 ft² of membrane area. The BW30-365 membrane carries a 34-mil feed spacer while the BW30-400 carries a 28-mil spacer. The BW30-365 is recommended for more fouling-prone feed waters because of the thicker feed spacer. Dow Water and Process Solutions also offers the BW30-400/34*i*, a 400ft² module with 34-mil spacer and the *i*LEC end caps. This module would be used as a replacement for the BW30-365 when a lower flux is required on fouling-prone feed waters, or in a new system to reduce the size of the system versus using the BW30-365 and still get the resistance to fouling offered by the thicker feed spacer.

Having selected the appropriate water flux and membrane type, the designer can use equation 10.1 to back calculate how many membrane modules are required to meet the desired productivity.

$$J_w = F_p * 1/MA * 1/N \quad (10.1)$$

where:

- J_w = water flux, gfd
- F_p = product flow rate, gallons/day
- MA = membrane area per module
- N = number of modules

Solving equation 10.1 for the number of modules, N , can give the designer an idea of what the array might look like. For a two-stage, 75% recovery system with a tapered design, 2/3 of the total number of modules would be in the first stage and the remaining 1/3 would be in the second stage. Once the array has been estimated, the design can be optimized with respect to other design variables including flow rates, scaling indexes, recovery, and Beta.

At this point continuing to create a design by hand calculation is very tedious considering all the variables that must be addressed by the design. Thus, design software has been developed to aid the designer in developing a design of an RO system. Some membrane manufacturers have made available to the public, design software that is specific for their membranes. Each software package, while different in presentation, delivers the same result: design of the RO unit, including array, operating pressure, scaling indices, and product and concentrate water qualities.

The software is based on stabilized, nominal performance of the membrane selected under design conditions. Actual performance may vary up to +/-15% of nominal, according to the Dow Water and Process Solutions-FilmTec Technical Manual.¹ The software will give warnings when the basic element operating parameters are exceeded, such as high element recovery or low concentrate flow. However, the software will not give warnings when the overall design is not practical. In fact, the Dow Water and Process Solutions FilmTec Technical Manual states:

"ROSA [Reverse Osmosis System Analysis] 6 only projects reverse osmosis and nanofiltration system performance from a User-controlled set of data input and design decisions. *The*

*program does not judge whether a system design is realistic [emphasis added] or optimized for a given set of conditions. It is the User's responsibility to review and judge the system design based on the anticipated or existing pretreatment, reasonable design guidelines, and experience."*¹

Hence, experience and common sense are necessary to ensure the design selected using the any design software package is realistic, particularly for feed streams other than well water with SDI < 3.

Table 10.1 lists four major U.S. membrane manufacturers and their respective design programs that are available to the public. Each design program is discussed in detail below. The programs can be downloaded from the manufacturers' respective websites. TriSep Corporation, Goleta, California also offers a design program that can be downloaded from their website; it is not covered in this chapter as TriSep is not generally considered a major US manufacturer. The other major U.S. manufacturer, GE Water & Process Technologies (Trevose, Pennsylvania), does not offer a design program that is available to the public. Applied Membranes®, Inc., (Vista, California), another U.S. manufacturer, does not offer a design program to the public.

Table 10.1 Four major US suppliers of RO membranes and their design programs Available to the public via download from their respective websites.

Membrane Manufacturer	Design Software Package
Dow Water Solutions--FilmTec (Minneapolis, MN) www.dow.com	Reverse Osmosis System Analysis (ROSA)
Toray Membrane, USA (Poway, CA) www.toraywater.com	Toray Design System (TorayDS)
Hydranautics (Oceanside, CA) www.membranes.com	Integrated Membrane Solutions™ (IMSDesign)
Koch Membrane Systems (Wilmington, MA) www.kochmembrane.com	ROPRO®

The selection of whose software program to use depends entirely on which membrane manufacturer is specified by the customer. Each RO system designer may have a favorite program that they use to provide projection information should the membranes of choice not be specified. In most cases, it makes sense to run several programs and compare/contrast the differences among them to find which membrane performance meets the requirements of the specific application.

While each program is unique to its particular manufacturer's membranes, there are similarities among the programs. The details, including similarities and differences among the four design programs listed in Table 10.1, are discussed below.

10.1 ROSA Version 6.1

Dow Water and Process Solutions-FilmTec's ROSA program version 6.1 has four input pages and one report page, each tabbed on the bottom of the screen. The five tabs are:

- Project Info
- Feed Data
- Scaling
- System Configuration
- Report

The Project Info screen allows the designer to input information about the project. This screen also lets the designer input different cases, with each case having its own set of design information. This allows the designer to examine several variations in design without having to input the bulk of the data again and again.

The feed screen is where the influent water quality is entered. ROSA allows the designer to enter more than one feed stream and have them blended in whatever combination is required. This screen also has an input for the type of water being treated based on SDI and source. This is an important input as it affects the fouling rate of the membrane and should agree with the overall average flux output (see Chapter 9.1.1). The program considers this input when issuing warnings on concentrate flow rates that are too low.

The scaling screen provides the designer with feed and concentrate scaling indices for the recovery and temperature that the designer inputs. This screen also allows the designer to select acid

or caustic feed for pH adjustment or to select sodium softening to reduce hardness. ROSA will calculate the softened feed water quality, should this option be selected.

System design variables are input in the configuration screen. This screen is shown in Figure 10.1. Several of the inputs are highlighted in Figure 10.1. These inputs are described below:

1. The number of passes is entered in this input box. ROSA allows for a maximum of two passes.
2. Permeate flow rate is entered here. Units are selected on the project info screen.
3. System recovery is entered here.
4. The number of stages in the pass and which stage is currently being configured is entered in these two input boxes. ROSA allows for an unlimited number of stage per pass.
5. Permeate backpressure per stage (if any) is entered. The back pressure will determine the effluent water quality and Beta, as discussed in Chapter 9.6.
6. The system array is determined at this point. The designer enters the number of pressure vessel per stage and the number of membrane elements or modules per pressure vessel. The program then calculates the number of elements in the stage.
7. The type of membrane is selected here. ROSA offers all membrane products, from tap water elements through seawater and nanofiltration membranes. Specifications for each membrane are available by clicking on the Specs button. ROSA also allows the use of difference elements in the same pass so that each stage can have difference elements (see Chapter 17.2.20).

Other features on the configuration screen include the following:

1. **Recirculation Loops:** Allows the designer to either blend permeate or recycle concentrate to the feed of that specific pass, or recycle concentrate from a second pass to the feed to the first pass.
2. **Fouling Factor:** The designer can input the rate of fouling of the membrane based on the feed water quality. A fouling factor of 1.00 is used for new membranes. For

Figure 10.1 Configuration input screen in Dow Water and Process Solutions ROSA design program.

brackish water membranes, a fouling factor of 0.85 is a good approximation for loss of pressure due to compaction and irreversible fouling at year 3 of membrane life. (the fouling factor 0.85 corresponds to a 15% loss in flux due to compaction and irreversible fouling). Note that if the feed water source is low quality (i.e., high in suspended solids) than a fouling factor greater than 0.85 (actual fouling factor number is less than 0.85) should be used. For example, treating municipal effluent the designer may want to select a fouling factor of 0.75. Appropriate estimation of the fouling factor is important to properly size the high-pressure feed pump and motor for the RO system.

3. **Feed Pressure:** This value can be input by the designer or will be calculated by the program. Feed pressure is typically input only when running projections for existing systems.

Once all the input screens have been completely populated, selecting the "Report" tab will allow the program to perform the

design calculations. Figures 10.2 through 10.6 show the output from ROSA. These figures are discussed below.

Figure 10.2 shows the primary design output from ROSA. The primary design output includes design inputs as well as calculated outputs. Inputs included are:

- Fouling factor
- Water classification and SDI
- Recovery
- Membrane element selected
- Array

Calculated outputs include the following:

- Feed pressure: This is the projected pressure required to the first stage of the RO system. The pressure is a function of the flow rate, water temperature, array, and fouling factor selected.
- Average pass flux: This is the projected average flux over the entire pass, including all stages. The average flux, along with the water classification and SDI, should agree with the recommendations listed in Table 9.1.
- Feed flow: This is the projected feed flow per stage. The designer should divide the calculated feed flow per stage by the number of pressure vessels in that stage. The result should fall under the recommendations for feed flow rate per pressure vessel, Chapter 9.1.1, and Table 9.2.
- Concentrate flow: This is the projected concentrate flow per stage. The designer should divide the calculated concentrate flow per stage by the number of pressure vessels in that stage. The result should agree with the recommendations given in Table 9.3 for the minimum concentrate flow rate per pressure vessel.
- Concentrate pressure: Concentrate pressure can be used to determine the projected pressure drop through the RO system. Projected feed pressure minus the projected first stage concentrate pressure is the projected pressure drop over the first stage. The concentrate pressure from the first stage minus that for the second stage is the projected pressure drop over the second stage.

Project Information:													
System Details													
Feed Flow to Stage 1			438.16 gpm		Pass 1 Permeate Flow			332.95 gpm		Osmotic Pressure:			
Raw Water Flow to System			438.16 gpm		Pass 1 Recovery			75.99 %		Feed		1.33 psig	
Feed Pressure			201.88 psig		Feed Temperature			60.1 F		Concentrate		5.36 psig	
Fouling Factor			0.85		Feed TDS			179.32 mg/l		Average		3.35 psig	
Chem. Dose (100% H2SO4)			0.00 mg/l		Number of Elements			90		Average NDP		180.52 psig	
Total Active Area			32850.00 ft²		Average Pass 1 Flux			14.60 gfd		Power		48.11 kW	
Water Classification: RO Permeate SDI < 1										Specific Energy		2.41 kWh/kgal	
Stage	Element	#PV	#Ele	Feed Flow (gpm)	Feed Press (psig)	Recirc Flow (gpm)	Conc Flow (gpm)	Conc Press (psig)	Perm Flow (gpm)	Avg Flux (gfd)	Perm Press (psig)	Boost Press (psig)	Perm TDS (mg/l)
1	BW30-365	10	6	438.16	196.88	0.00	208.83	182.36	229.33	15.08	0.00	0.00	0.75
2	BW30-365	5	6	208.83	177.36	0.00	105.21	163.55	103.62	13.63	0.00	0.00	1.60

Figure 10.2 Primary design output of Dow Water and Process Solutions ROSA design program.

And so forth. When running a projection for an existing system, the projected and actual pressure drop data should be contrasted and used to determine whether fouling or scaling may be occurring.

The ROSA output continues in Figure 10.3. This figure shows the actual and projected water quality throughout the RO system. The “Feed” column represents the raw feed water to the system. “Adjusted Feed” is the projected water quality after pH-adjust or after softening. In this case, “Stage 2 Concentrate” is the project quality of the overall concentrate from the RO. If 3 stages were used, then the “Stage 3 Concentrate” would be the project quality of the overall concentrate from the RO. “Permeate Total” is the projected overall permeate concentration from the system.

Figure 10.4 continues the ROSA output showing the projected performance of each element in the RO system. This output includes the recovery per element, which should be limited to no more than 15%. Higher recovery per element will trigger a design warning.

Design and solubility warnings are shown in Figure 10.5. The design warnings cover element operating parameters such as recovery per module, but do not cover the feasibility of the design as selected by the designer. Solubility warnings cover the LSI (see Chapter 3.9) and specific scale-forming compounds such as barium sulfate.

Figure 10.6 shows the projected scaling indices for raw feed, adjusted feed, and concentrate streams. The output includes LSI, alkalinity,

Pass Streams (mg/l as ion)							
Name	Feed	Adjusted Feed	Concentrate		Permeate		
			Stage 1	Stage 2	Stage 1	Stage 2	Total
NH4	0.00	0.00	0.00	0.00	0.00	0.00	0.00
K	2.50	2.50	5.19	10.19	0.05	0.12	0.07
Na	21.00	22.91	47.99	95.12	0.08	0.19	0.11
Mg	5.90	5.90	12.37	24.54	0.01	0.03	0.02
Ca	17.00	17.00	35.64	70.72	0.03	0.07	0.04
Sr	0.08	0.08	0.17	0.33	0.00	0.00	0.00
Ba	0.07	0.07	0.15	0.29	0.00	0.00	0.00
CO3	0.57	0.57	1.96	5.47	0.00	0.00	0.00
HCO3	89.07	89.07	185.05	363.71	0.44	0.83	0.56
NO3	3.00	3.00	6.21	12.15	0.08	0.19	0.11
Cl	19.00	19.00	39.84	79.03	0.03	0.08	0.05
F	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO4	16.00	16.00	33.56	66.60	0.02	0.04	0.03
Boron	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SiO2	3.20	3.20	6.70	13.26	0.02	0.04	0.02
CO2	1.17	1.17	1.61	2.47	1.17	1.74	1.35
TDS	177.39	179.32	374.84	741.46	0.75	1.59	1.01
pH	8.10	8.10	8.25	8.33	5.87	5.97	5.91

Figure 10.3 Actual feed and projected water quality output of Dow Water and Process Solutions ROSA design program.

Stage Details						
Stage 1 Element Recovery		Perm Flow (gpm)	Perm TDS (mg/l)	Feed Flow (gpm)	Feed TDS (mg/l)	Feed Press (psig)
1	0.09	3.94	0.56	43.82	179.32	197.31
2	0.10	3.88	0.61	39.87	196.95	193.77
3	0.11	3.83	0.68	35.99	218.07	190.72
4	0.12	3.79	0.77	32.17	243.89	188.12
5	0.13	3.76	0.88	28.38	276.27	185.96
6	0.15	3.74	1.03	24.62	318.22	184.19
Stage 2 Element Recovery		Perm Flow (gpm)	Perm TDS (mg/l)	Feed Flow (gpm)	Feed TDS (mg/l)	Feed Press (psig)
1	0.09	3.62	1.13	41.76	374.84	177.78
2	0.09	3.55	1.26	38.14	410.18	174.48
3	0.10	3.49	1.43	34.59	452.02	171.61
4	0.11	3.42	1.63	31.11	502.37	169.16
5	0.12	3.36	1.91	27.68	564.12	167.09
6	0.14	3.30	2.27	24.32	641.57	165.37

Figure 10.4 Element-by-element performance projection output of Dow Water and Process Solutions ROSA design program.

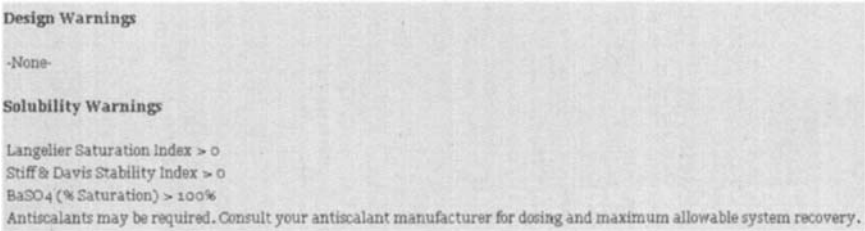


Figure 10.5 Design and solubility warnings output of Dow Water and Process Solutions ROSA design program.

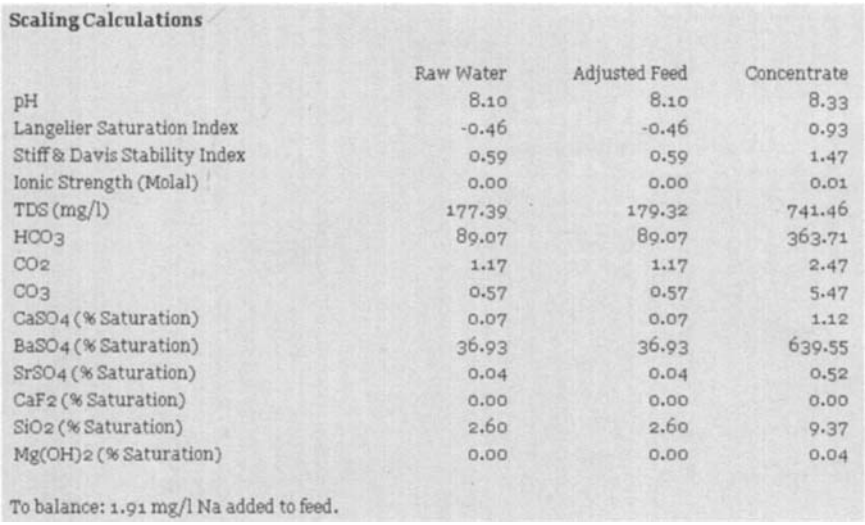


Figure 10.6 Projected scaling indexes of the Dow Water and Process Solutions ROSA design program.

and the percent saturation for several salts including barium sulfate, calcium fluoride, and silica.

The printed output for ROSA is three pages. The format is the same as the output screen.

At the bottom of the output screen is a link to an “Overview Report.” This report page provides a simplified process flow diagram, as shown in Figure 10.7. The report also provides an overview of the system configuration as well as the flow rates for each stream depicted in Figure 10.7. This Overview Report can also be printed out and used as a cover sheet for the main output page (Figures 10.2 – 10.5).

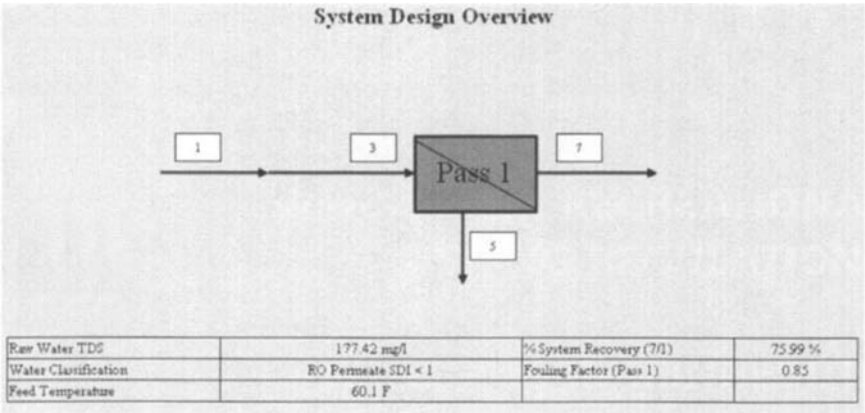


Figure 10.7 Dow Water and Process Solutions ROSA Overview Report.

10.2 TorayDS Version 1.1.44

Toray Membrane USA’s “TorayDS” reverse osmosis design software has a lot of information packed into first screen upon start-up. “TEACHMODE” gives a step-by-step guide as to what data to enter when and how. Further, the basic inputs are highlighted in red, so that the designer can easily identify them. Finally, the designer can elect to have a “rudimentary” array and/or membrane selected by the program itself, by clicking on the “Auto – Design” or “Auto – Element” buttons, respectively.

The program has one main input screen, plus one screen for water quality inputs. Figure 10.8 shows the main input screen that comes up when entering the program. Rather than a tabbed approach such as ROSA, Toray uses a simple diagram to enter design parameters. The terminology is a bit different using this program. This will be explained as encountered in the specific descriptions below (note that the numbers refer to the corresponding inputs in the ROSA program).

1. “Select 2-Pass”—this is where the number of passes is input. Default is one pass. Clicking on this button will allow for the second pass screen to load, after information for the first pass is entered.
2. The permeate flow rate is entered here. Clicking of the “Permeate” button will open up a box into which the

- permeate flow is entered. The minimum permeate flow rate is shown in the box as a guide. Units are selected using the "Options" menu at the upper left of the screen.
3. System recovery is entered here. Clicking on "Recovery" will open up a box into which the recovery can be entered. Recovery upper and lower limits are shown in the box as guides.
 4. The number of banks (or stages) is entered here. Clicking on the right arrow increases the number of stages, while the left arrow decreases the number of stages.
 5. This is where the permeate backpressure per stage is entered. Clicking on the control valve will open up a box into which the backpressure is entered. Minimum and maximum backpressures are given as guides.
 - 5.a. Individual stage backpressures can be entered here. This would be used, for example, reduce the recovery for the selected stage and thereby increase the feed flow rate to the subsequent stages.
 - 5.b. Total permeate backpressure is entered here. This is used when permeate from each stage is being sent to a common unit operation or tank which exerts a common backpressure on the system.
 6. This is where the system array is entered. Use the Up or Down arrows to increase or decrease the number of elements per stage and the number of pressure vessels per stage, respectively.
 7. The membrane elements are selected here. Double clicking on the input box will open up a window containing all Toray membrane elements and specifications. The designer can select a membrane or use the "AutoSelect" button and the program will select an appropriate model. Each stage can contain a different type of membrane, if required (see Chapter 17.2.20).

Other data that should be entered into the screen shown in Figure 10.8 includes the "Water Type." Clicking on this button will open up a box that allows the designer to select the feed water source (see Chapter 9.1 for details about the importance of the source of the feed water to the RO system). This box also lists the recommended percent salt passage increase per year and fouling factor that corresponds to the type of feed water selected. (These guidelines should be entered

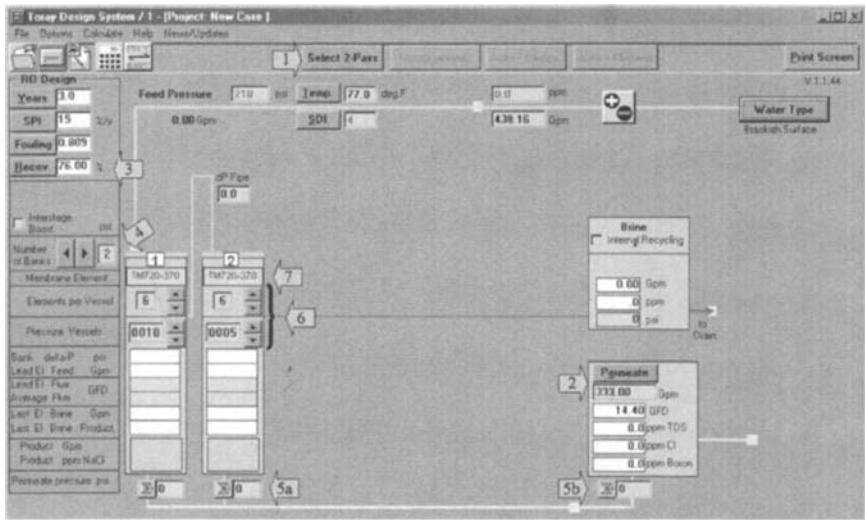


Figure 10.8 Primary input screen of TorayDS design program.

into the “SPI (salt passage increase)” and “Fouling” buttons on the primary input screen.) The default values are loaded for the purposes of generating warnings unless the designer over-rides them (they should only be changed by experienced designers). Clicking on the “Design Guidelines” button reveals most of the important guidelines for RO system design as functions of water type. Guidelines include recommended minimum and maximum flow rates per pressure vessel, recommended flux rates, recommended maximum differential pressure per element and per pressure vessel, and so on. Note that, “...design guideline fouling factors are estimates only, based on historical data from similar water sources. The actual fouling rate is entirely dependent on site conditions.”¹ Once the water type has been completed, warnings will appear on the primary input screen either in a box or as a red number in the design inputs if any of the recommended guidelines have been exceeded by the selected design.

After items one through seven above plus the water source are entered into the screen shown in Figure 10.8, the feed water analysis is entered. The input screen can be accessed from the menu under “Options,” or by clicking on the “+/-” button on the input screen (located just left of the “Water Type” button). The designer can enter water quality data as either ppm as ion, ppm as calcium carbonate, milli-equivalents per liter or milli-moles per cubic meter.

The program calculates total dissolved solids (TDS), conductivity, and the Langelier and Stiff-Davis Saturation Indexes (see Chapter 3.9). The screen allows the designer to use sodium, calcium, magnesium, chloride, sulfate, or bicarbonate to balance the water analysis. Water quality for up to five feed streams can be entered and blended together to make the total, combined feed water to the system.

When the designer has completed water quality entries, clicking on the “Go to Pretreatment” button will bring up a screen showing the scaling indexes. This screen allows the designer to select pretreatment with acid or caustic, sodium softening, or weak acid cation resin for calcium scale control. The use of an antiscalant (SHMP—see chapter 8.2.3—is the only specific antiscalant that can be selected) can be added for sulfate-scale control. Finally, degasification or caustic addition can be selected for reducing carbon dioxide in the RO feed water.

Once the pretreatment screen has been completed, options to return to the water analysis screen or the RO design screen are available to the designer. Clicking on the “RO – Design” button will return the designer to the primary input screen, Figure 10.9. If all inputs are in order, the “inputs Missing” button in Figure 10.8 is replaced with the “Calculate or F9” button. Clicking on this button or the “Calculate” menu will run the projection.

The RO design screen shown in Figure 10.10 appears when the calculation is completed. All outputs are populated including performance of each stage of the system as well as the permeate quality and concentrate (“brine”) pressure.

Clicking on the view and print results button (with the “eye” and sheet of paper icon in the upper left under the menu) will bring up complete performance data, including flows and design inputs, water quality, and disclaimers. The standard printout is about two to three pages, including the disclaimers. The designer can select data to be displayed with one or two decimal points. Two decimal points are useful when considering two-pass RO systems. A process flow diagram (PFD) is available as well. Note that the permeate backpressure information is only displayed on the PFD, not in the text data.

10.3 Hydranautics IMS Design Version 2008

The Hydranautics design software package offers ultrafiltration (UF), RO, and UF+RO design options. The RO only option has two input

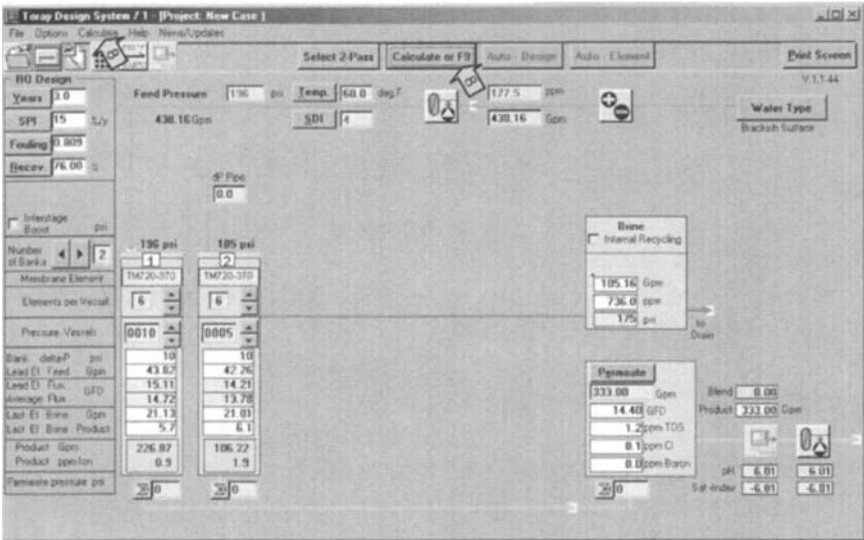


Figure 10.9 Primary input screen showing “calculate” button for TorayDS design program.

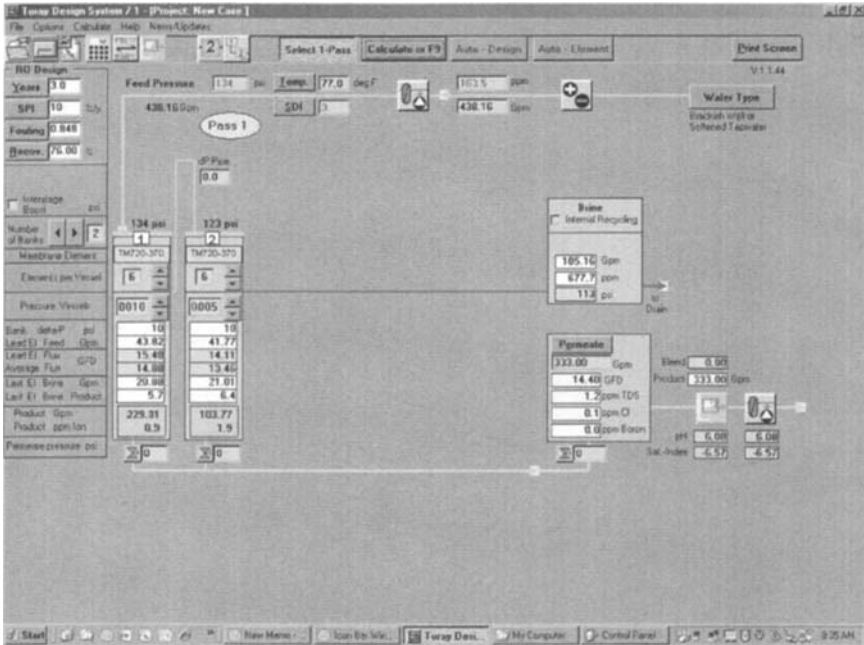


Figure 10.10 Design output screen for TorayDS design program.

screens, “Analysis” and “File.” The menu bar allows the designer to go to the “Analysis” or “File” screens or to the help screen.

The help menu opens up a long list of issues that the designer can find assistance with, as shown in Figure 10.11. Additionally, several of the program inputs allow the designer to double click on the input query and information about that issue will pop up on the screen.

The first input screen is the water analysis screen, shown in Figure 10.12. This screen is where the designer inputs the water analysis, either as ppm ion or ppm calcium carbonate. The screen also has inputs for iron, SDI, hydrogen sulfide, and turbidity. There is a drop-down menu where the feed water source is input. The bottom of the page lists the scaling indices.

Once the “analysis” page has been completed, the “RO Design” input screen can be called up by clicking on the appropriate button. This screen is shown in Figure 10.13. The numbered inputs correspond to the input numbers on the ROSA primary input screen (Figure 10.1):

1. Number of passes,
2. Permeate flow rate,
3. Recovery,

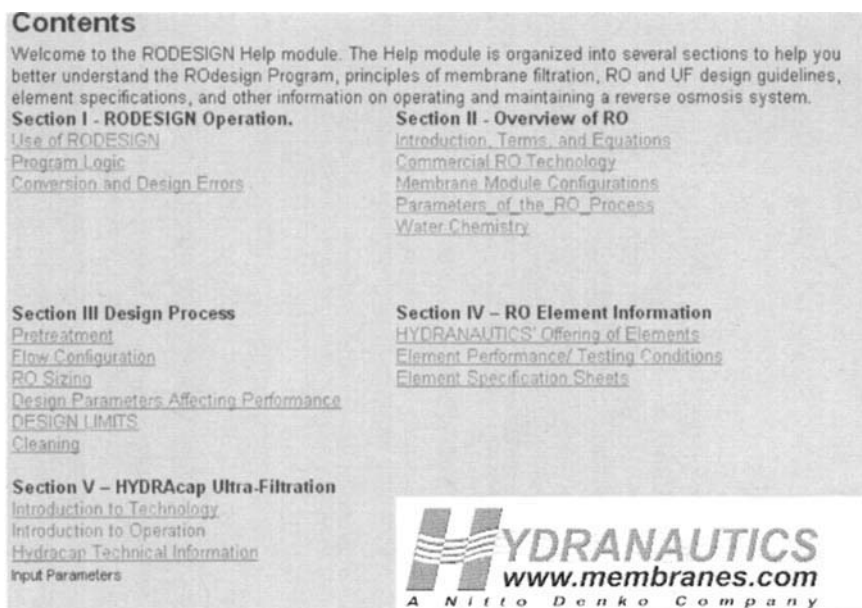


Figure 10.11 “Help” contents of Hydranautics IMS Design program.

Project	Temple Inland PaperBoard		Code	TEMPLE	Feed	Well Water		Date		
pH	8.10		Turb	0.5	E cond	269	uS/cm	CO2	1.200 ppm	
Temp	25.0 C		SDI	4.0 15min	H2S	0.0	ppm	Fe	0.000 ppm	
Ca	17.0	ppm		0.85	meq	CO3	0.7	ppm	0.02 meq	
Mg	5.9	ppm		0.49	meq	HCO3	73.0	CaCO3	1.46 meq	
Na	22.8	ppm		0.99	meq	SO4	17.0	CaCO3	0.34 meq	
K	2.5	ppm		0.06	meq	Cl	19.0	ppm	0.54 meq	
NH4	0.0	ppm		0.00	meq	F	0.0	ppm	0.00 meq	
Ba	0.072	ppm		0.00	meq	NO3	2.5	CaCO3	0.05 meq	
Sr	0.079	ppm		0.00	meq	B	0.05	ppm	0.00 meq	
						SiO2	3.2	ppm	0.00 meq	
Total positive				2.39 meq	Autobalance		Total negative			2.41 meq
Calculated TDS	180		ppm	Ionic strength	0.003		Print			
CaSO4 saturation	0.3		%	BaSO4 saturation	117.6		%			
Silica saturation	2.3		%	SrSO4 saturation	0.1		%			
Saturation Index	-0.3		Langelier	Osmotic pressure	1.5		psi			
							Save			

Figure 10.12 Water analysis input screen of Hydranautics IMS Design program.

Project	Temple Inland PaperBoard		Code	TEMPLE	Feed	Well Water		Date		
pH	8.10		Turb	0.5	E cond	269	uS/cm	CO2	1.200 ppm	
Temp	25.0 C		SDI	4.0 15min	H2S	0.0	ppm	Fe	0.000 ppm	
Ca	17.0	ppm		0.85	meq	CO3	0.7	ppm	0.02 meq	
Mg	5.9	ppm		0.49	meq	HCO3	73.0	CaCO3	1.46 meq	
Na	22.8	ppm		0.99	meq	SO4	17.0	CaCO3	0.34 meq	
K	2.5	ppm		0.06	meq	Cl	19.0	ppm	0.54 meq	
NH4	0.0	ppm		0.00	meq	F	0.0	ppm	0.00 meq	
Ba	0.072	ppm		0.00	meq	NO3	2.5	CaCO3	0.05 meq	
Sr	0.079	ppm		0.00	meq	B	0.05	ppm	0.00 meq	
						SiO2	3.2	ppm	0.00 meq	
Total positive				2.39 meq	Autobalance		Total negative			2.41 meq
Calculated TDS	180		ppm	Ionic strength	0.003		Print			
CaSO4 saturation	0.3		%	BaSO4 saturation	117.6		%			
Silica saturation	2.3		%	SrSO4 saturation	0.1		%			
Saturation Index	-0.3		Langelier	Osmotic pressure	1.5		psi			
							Save			

Figure 10.13 Primary design input screen of Hydranautics IMS Design program.

- 4. Number of stage per pass,
- 5. Permeate backpressure,
- 6. Array,
- 7. Membrane selection.
- 8. Run—perform the design calculations.

Other inputs into this screen include the following:

- pH and chemical feed are located at the upper right and upper left of the screen, respectively. The designer can select sulfuric acid or caustic (or none) for chemical feed. Once the chemical feed is selected, the desired pH is entered. The chemical dosing rate required will appear in the in the top center of the screen.
- Membrane age. For general projections, 3 years should be selected which assumes a 3-year membrane life. This input works closely with the flux decline and salt passage increase inputs discussed below. Selecting performance at end-of-life for the membranes will yield the operating parameters necessary after years of fouling and scaling of the membrane.
- Flux decline per year input is located in the upper left of the screen. Recommended percentage flux decline per year as a function of feed water quality is shown in Table 10.2. Note that the flux decline increases as the water quality gets worse.
- Salt passage increase is located in the upper left of the screen, just below the flux decline input. Recommended percentage salt passage increase per year as a function of feed water quality is listed in Table 10.3. Note that the salt passage increases as the feed water quality goes down.
- Clicking on the “Recalc Array” button will change the selected array to one that is more appropriate for the conditions entered into the program.

Table 10.2 Hydranautics’ recommended flux decline percentage per year as a function of feed water quality.

Flux Decline per Year (%)	RO Permeate	Brackish Well Water	Brackish Surface Water	Tertiary Wastewater
Conservative	7	10	10	18
Maximum	3	7	7	7

Table 10.3 Hydranautics’ recommended salt passage increase percentages per year as a function of feed water quality.

Salt Passage Increase per year (%)	RO Permeate	Brackish Well Water	Brackish Surface Water	Tertiary Wastewater
Conservative	7	15	15	15
Minimum	5	10	10	10

- “Auto Display” allows the designer to opt for a full report rather than a summary report of the projection. The full report that comes up on the screen is what the hard copy will look like.

The summary report that comes up after the calculations are performed is shown in Figure 10.14. The top half of the screen shows the input values, while the bottom half gives a summary of the projected performance. To change any variable and rerun the

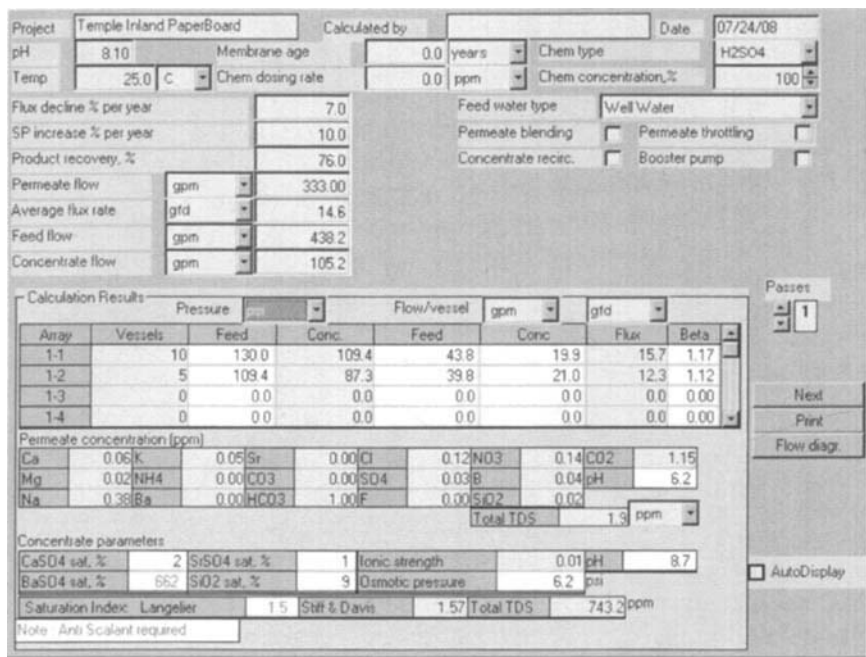


Figure 10.14 Summary report output screen of Hydranautics IMS Design program.

projections, click on “Next.” This brings the designer back to the screen shown in figure 10.13.

The hard copy output from the Hydranautics design program is two pages. The first page of the hard copy output shows the inputs and project outputs for the design, the projected water quality, and the scaling indices. The second page includes the element-by-element projection data, including Beta values.

10.4 Koch Membranes ROPRO Version 7.0

Koch Membrane Systems ROPRO design program has four input screens that are accessed via buttons or the menu at the top left of the main screen. The four input screens are:

- Project Information
- Feed Analysis
- Flow Rates Recovery
- Array Specification

Each screen has a “Help” button that provides information about the inputs on that screen. Furthermore, there is an “Input Wizard” (accessed under menu “Tools—Inputs—Input Wizard,” or by the button with the magician’s hand and wand) that walks the designer through the design input screens as follows:

- Project information
- Unit selection
- Water analysis
- pH adjustment
- Flow rates and recovery
- Array configuration

Figure 10.15 shows the main screen that opens when the program is initiated. This screen does not have any actual inputs other than the “Go” or run program icon at the top center of the screen (see call out number “8” in Figure 10.1). The screen shows a process flow diagram including the RO membranes and feed pump, chemical feed pretreatment (if needed), and post-treatment degasification and chemical treatment (if needed). The program returns to this screen after each of the other input screens are completed.

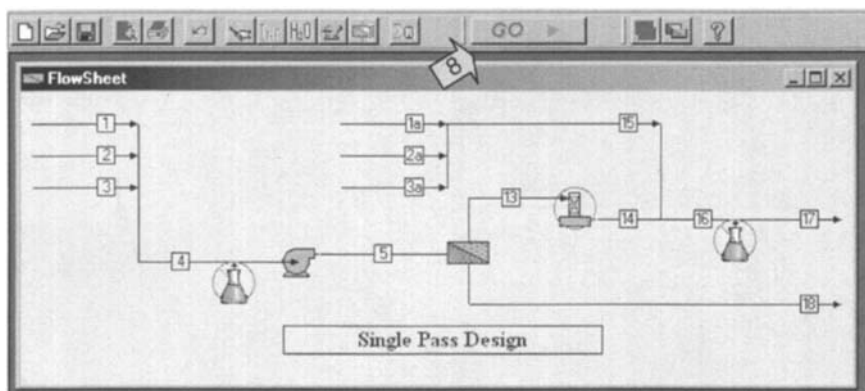


Figure 10.15 Primary input screen of the Koch ROPRO design program.

The Project Information screen is accessed by the menu under "Tools—Input," or the "INF" button at the top of the main screen. Project name, designer name, and the number of passes (along with recycle and booster pump, if required) are entered into this screen.

The Feed analysis screen is reached via the menu under "Tools—Input—Water Analysis," or by the "H₂O" button at the top of the main screen. Up to three feed analyses can be entered. Concentration can be entered as ppm ion, ppm as calcium carbonate, or as milliequivalence per liter. Standard species, such as calcium and sodium, are entered. There is also space for four designer-defined species of interest. Examples of designer-defined species include total organic carbon, boron, and the like. Note that the program will not automatically calculate permeate and concentrate concentrations for these designer-defined species.

Flow rates and recovery are entered into the screen that is accessed under the menu, "Tools—Input—Flow Rates & Recovery," or the button with the "pump" icon. Figure 10.16 shows this input screen. This screen allows the designer to select the mix among three potential feed streams in addition to the product (or feed or concentrate) flow rate and recovery (noted as items 2 and 3, respectively). Fouling allowance is also entered in this screen.

Figure 10.17 shows the membrane array configuration screen. This screen is accessed via the menu, "Tools—Input—Array Specification," or the button with the "RO membrane" icon. The designer can select up to 5 stages (called "banks") per pass

(item 4). Item 6 indicates where the array is entered. Permeate back-pressure is entered at item 5. The membrane module type is entered at item 7. Note that by clicking the box to the left of each of inputs 5 – 7, the same input carries over to all stages of the RO pass. Selecting the “array Wizard” button will enable the program to select the best array for the given inputs.

Once all inputs are entered, the designer should run the calculations by either selecting the button, “GO,” shown in Figure 10.15, or by using the menu, “Tools—Calculate.”

Figure 10.18 shows the output page. This screen shows the input variables such as the selected membrane, permeate flow rates, and recovery. Calculated outputs include:

- Operating pressure
- Permeate quality (TDS)
- Flux
- Beta
- Differential pressure
- And saturation indexes

Problem outputs are noted in red. Clicking on the “Warnings – Click to see Preview” button located in the upper right side of the

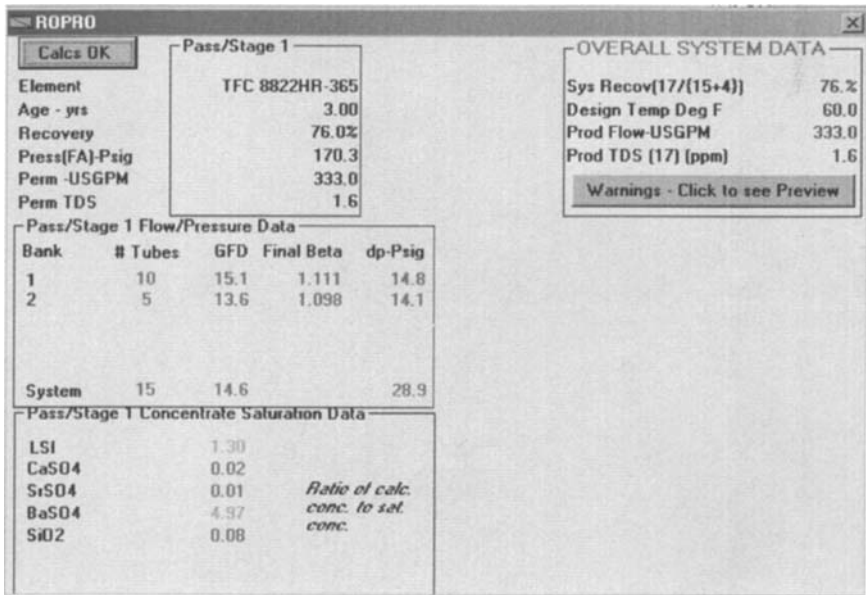


Figure 10.18 Output screen of the Koch ROPRO design program.

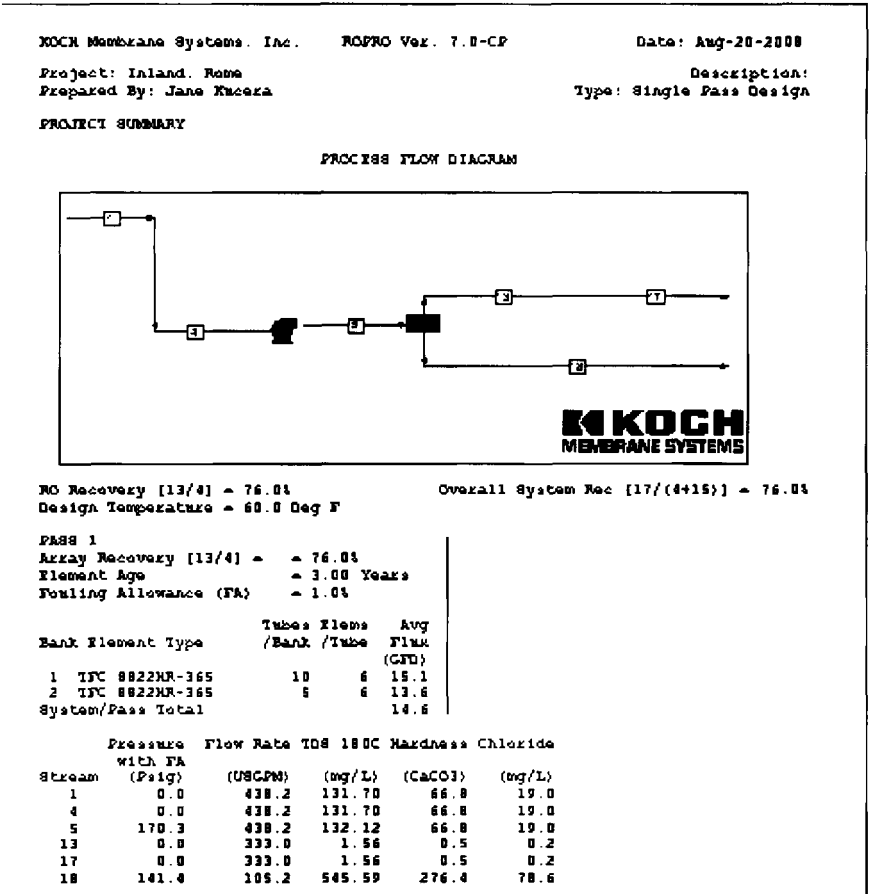


Figure 10.19 A portion of the printed output of Koch ROPRO design program.

screen will bring up a preview of the printed output, a portion of which is shown in Figure 10.19. The printed output is brief and includes warnings and disclaimers.

Reference

- 1. Madalyn Epple, Toray Membrane USA, personal communication, August 15, 2008.

4

OPERATIONS

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On-Line Operations*

The keys to understanding how well an RO system is operating are to take the proper data, to use the appropriate analytical techniques to interpret these data, and to perform maintenance to keep the RO and pretreatment systems operating as designed. This chapter covers both data that should be collected and how to interpret them, as well as a preventive maintenance schedule to help keep the RO and pretreatment systems performing well.

11.1 Reverse Osmosis Performance Monitoring

Performance of “conventional” deionization technologies, such as ion exchange, is well known and understood. In the case of RO, however, operator and engineers do not have a long history with this technology, and many lack the understanding of performance that comes with experience. Furthermore, monitoring of an RO system is not straightforward; observed performance is not always indicative of what is actually happening at, on, or in the membrane.

11.2 Data Collection

Table 11.1 lists primary data points that should be monitored to determine how an RO system is basically functioning.¹ The parameters listed in Table 11.1 all affect the product flow rate and/or product quality, as described in Chapter 9. Of all the data points identified in Table 11.1, feed, product, and reject data are most commonly measured, and most RO skids are constructed with required instrumentation for measuring these variables.

*Adapted from “Reverse Osmosis Performance: Data Collection and Interpretation,” originally presented at the 28th Annual Electric Utility Chemistry Workshop, Champaign, IL, May 2008, and published in “UltraPure Water® Journal, www.ultrapurewater.com, April, 2009.

Table 11.1 Primary monitoring variables for an RO system.

	Feed	Interstage	Product	Reject
Pressure	X	X	X	X
Flow	X	X	X	
Conductivity	X	X	X	X
Temperature	X			

However, many RO skids are not equipped with interstage-measuring capabilities. Interstage data are necessary to obtain when considering performance, as these data can assist with determining what is occurring within each stage of the RO system. Without interstage data, any degradation in performance cannot be identified as occurring in the first stage of the RO (and thus, most likely related to fouling of the membrane) or occurring in the final stage of the RO (most commonly related to scaling of the membrane).

In addition to the data in Table 11.1, it is necessary to gather other information so that a complete analysis of performance can be conducted.²

- **Permeate backpressure:** Are there valves on the permeate line or unit operations (including tanks) that may exert backpressure on the membranes? Has this backpressure been taken into account in the performance projections?
- **pH:** What chemicals and concentrations are being added, what is the dosing set point, and how is dosage being controlled?
- **Oxidizing biocide:** Which one(s) is being used and how is its dosage controlled? Also, how is it being removed and how is that being controlled?
- **Reducing agent (Anti-oxidant):** Which one(s) is being used and what is the feed protocol? How is the dosage being controlled?
- **Non-Oxidizing biocide:** Which one(s) is being used and how is the dosage controlled? On-line or off-line usage?

- Antiscalant: What product is being used, what is the dosage, and how is the dosage controlled? Is flow proportional control being used?
- Other pretreatment chemical feeds: These include coagulants and flocculants. What is the nature of the chemical(s) being fed, the dosage, and how is the dosage controlled?
- Particle monitors: Particle monitors can be used to follow changes in the nature of the suspended solids in the feed water and throughout the pretreatment system. Do the monitors indicate a change in the number of particles of a given size? This could indicate a change in source water quality or in the efficiency of the pretreatment.

11.3 Data Analysis and Normalization

Parameters listed in Table 11.1 directly affect the observed product flow rate and observed salt passage through an RO membrane. Additional factors that affect membrane performance include the degree of membrane fouling, scaling, and degradation that has occurred. These three factors directly affect the observed product flow rate and observed salt passage, just as temperature, concentration, and pressure do. Because these operating conditions are constantly changing, it is not possible to compare the observed performance, such as product flow rate, at one time with the observed performance at another time.

11.3.1 Data Normalization

Data normalization was developed to allow for the direct comparison of performance at one time with that at another time by "neutralizing" via normalization the effects of temperature, pressure, and concentration on the performance parameter of interest. Normalization converts data collected at a given time and conditions to those at a designated baseline time, typically at start up of the RO system (or following installation of new membranes). In this manner, effects of temperature, pressure, and concentration are eliminated so that the only changes in normalized performance are due to membrane fouling, scaling, and/or degradation.

11.3.1.1 Normalized Product Flow

Equation 11.1 shows how temperature, in terms of the temperature correction factor, TCF; pressure, P ; and concentration, as osmotic pressure, Π , are used to normalize product flow rate.³

$$\text{Normalized flow} = \frac{[(AAP_s - \Delta\Pi_s)]TCF_s}{[(AAP_a - \Delta\Pi_a)]TCF_a} \bullet \text{Actual flow} \quad (11.1)$$

where

AAP = the average applied transmembrane pressure

$$= P_{\text{feed}} - \frac{\Delta P}{2} - P_{\text{perm}} \quad (11.2)$$

where

P_{feed} = applied feed pressure

ΔP = pressure *drop* from feed inlet to concentrate effluent

P_{perm} = permeate pressure

$\Delta\Pi_n$ = difference between the osmotic pressure on the membrane feed and permeate sides

TFC = temperature correction factor (membrane and manufacturer dependent)

"s" = subscript for "standard" conditions

"a" = subscript for "actual" conditions

In practice, data normalization is calculated using a spreadsheet or other of computer program. The best programs are integrated into a package that includes the hardware to actually capture the raw data. This eliminates the need to manually enter data. In general, systems that require manual data entry do not stand up to the test of time; operators will usually cease manually entering data within the first couple of months after start-up, and they are left with only observed data with which to analyze performance. As discussed previously, observed data are unreliable due to the effects of pressure, temperature, and concentration on product flow and salt rejection.

Once the normalized product flow (NPF) is calculated, it is best graphed to observe trends. Figure 11.1 shows various trends in the NPF. Curves with a positive slope are indicative of membrane

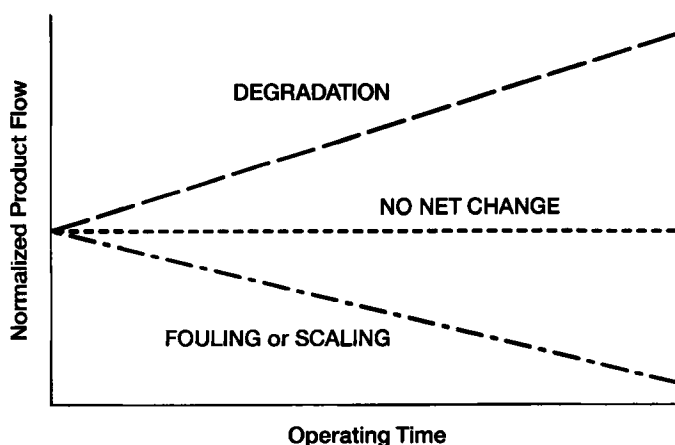


Figure 11.1 Trends in normalized product flow.

degradation, as more water is passing through the membrane at time $T > 0$ than at the initial operating conditions ($T = 0$). A negative slope is indicative of membrane fouling or scaling, as less water is passing through the membrane then would be at the initial operating conditions. A flat slope is indicative of no change in performance (this can mean no changes have occurred or that both degradation and fouling or scaling are occurring and, in effect, canceling each other out).

Case Study

Figure 11.2 shows the actual raw data from a facility operating on Delaware River Water.⁴ As the graph shows, the operators did an excellent job of keeping the productivity of the system steady at 340 gpm the design flow rate. Based on these raw data, one would believe that the RO system was operating well.

Figure 11.3 shows the normalized data for the same system.⁴ The normalized product flow exhibited an initial negative slope with time. Based on the information given in Figure 11.1, it appears as if the RO system was fouling or scaling, leading to a decrease in normalized permeate flow. Upon investigation, the operators were constantly increasing the operating pressure of the membrane system to force water through the increasingly thick layer of foulant or scale on top of the membrane. Hence, constant product flow rate was maintained despite fouling or scaling, due to the increase in operating pressure.

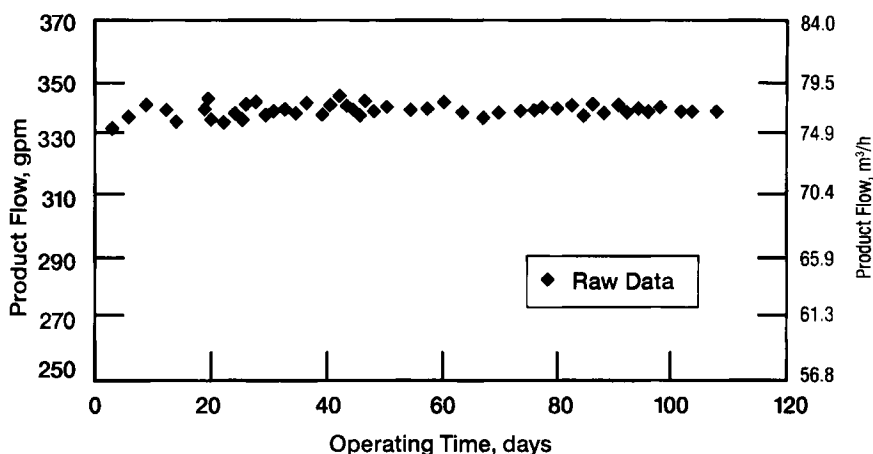


Figure 11.2 Actual product flow rate data from a facility operating on cold-lime softened Delaware River water.

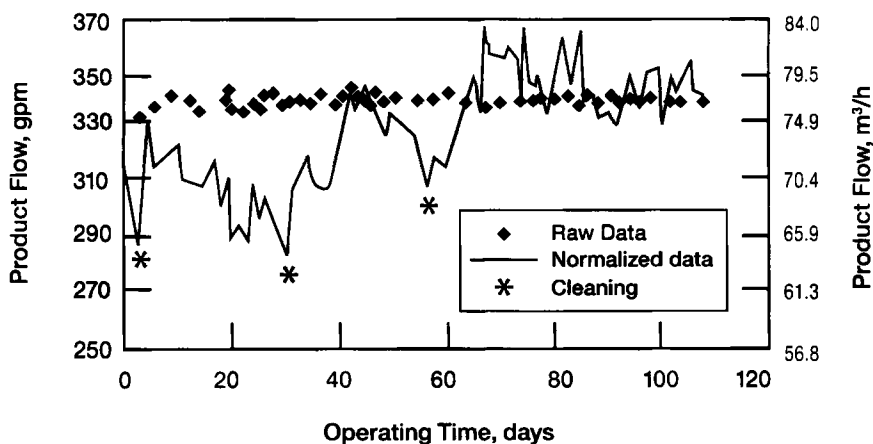


Figure 11.3 Normalize product flow rate data from a facility operating on cold lime softened Delaware River water.

Given that the operators were able to maintain a constant product flow rate, albeit by increasing the operating pressure, one might ask why we should even care about the NPF? When systems are allowed to operate in fouling or scaling mode for extended periods of time, the foulant or scale can become resistant to removal via cleaning. Thus, NPF is used to determine when it is time to clean the membranes before the surface contamination becomes permanent.

Further, continued operation under fouling or scaling conditions will result in the operating pressure reaching the maximum output allowed by the feed pump. If permitted to continue operating in this manner, the product flow rate would decrease as the pump could not supply higher pressure to overcome the resistance of the fouling layer. Hence, the key to keeping the membranes operating is to monitor the NPF and clean when the drop in NPF indicates it is time to do so. Cleaning of the membranes is recommended when the NPF drops 10% to 15% from initial operating conditions. Ideally, a cleaning should be scheduled at 10% drop and completed by the time the NPF drops 15% (see Chapter 13.2).

11.3.1.2 Normalized Salt Passage

Salt passage is another operating variable that is normalized. Since concentration is constantly changing, it is difficult to compare observed salt passage (or salt rejection) on a day-to-day basis. Normalizing the salt passage takes out the concentration and temperature variables, allowing the passage at any one time to be compared that at another time. Equation 11.3 shows how concentration affects the normalized salt passage through an RO membrane.³

$$\text{Normalized Salt Passage} = \left\{ \frac{\text{EPF}_a}{\text{EPF}_s} \cdot \frac{\text{CFC}_s}{\text{CFC}_a} \cdot \frac{\text{STCF}_a}{\text{STCF}_s} \cdot \frac{C_{fa}}{C_{fs}} \right\} \cdot \%SP_a \quad (11.3)$$

where

EPF = average permeate flow divided by the number of membrane modules

STCF = salt transport temperature correction factor (from membrane manufacturer)

Cf = feed salt concentration

%SP = percent salt passage

CFC = concentration of the feed-concentrate:

$$\text{CFC} = C_f \cdot \frac{\ln \frac{1}{1-Y}}{Y} \quad (11.4)$$

$$Y = \frac{\text{product flow}}{\text{feed flow}} \quad (11.5)$$

and

"s" = subscript for "standard" conditions

"a" = subscript for "actual" conditions

Monitoring salt passage is important in tracking membrane scaling and/or degradation. In both of these cases, salt passage will increase with time. If membrane degradation is occurring, this increase in salt passage is easy to understand. In the case of membrane scaling, however, the increase in salt passage is not so intuitive. Consider the following example.

Figure 11.4 depicts a cross-section of a membrane with a layer of calcium carbonate scale on the surface. The concentration of calcium at the membrane surface, Z , is higher than that in the bulk feed, X , since the concentration at the surface has reached saturation. The membrane passes salts based on what concentration is actually next to the membrane. In this case, the membrane is exposed to a saturated concentration, not the lower bulk solution concentration. Even though the percent passage of calcium through the membranes stays constant, the scaled membrane will yield higher permeate concentration of calcium. This is because the concentration of calcium that the membrane is exposed to at the membrane surface is higher than the bulk solution concentration of calcium, $[Z], [X]$.

Normalized salt passage is generally not used as the primary indicator of when to clean membranes. This is because normalized product flow and/or differential pressure drop (see below) will usually indicate problems with the membranes before product quality becomes an issue. However, normalized salt passage should be used in conjunction

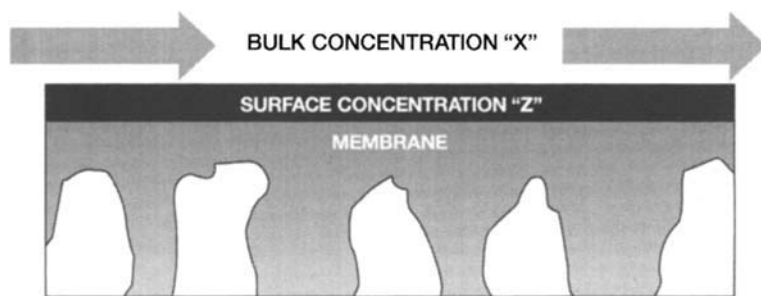


Figure 11.4 Cross-section of an RO membrane with a layer of scale on the surface of the membrane. Concentration of scale formers is higher at the membrane surface, Z than in the bulk solution, X .

with NPF and pressure drop to diagnose and troubleshoot problems with the RO system.

11.3.1.3 Normalized Pressure Drop

Pressure drop through the RO system can also be used as an indicator of when it is time to clean membranes. Pressure drop is a direct measure of the pressure loss due to friction caused by scale or foulants on the membrane or the feed spacer.

The energy lost from pressurized feed water is absorbed by the membrane module materials, which can cause the materials to shift within the module when the degree of fouling or scaling is severe. This can lead to telescoping of the membrane leaves, resulting in

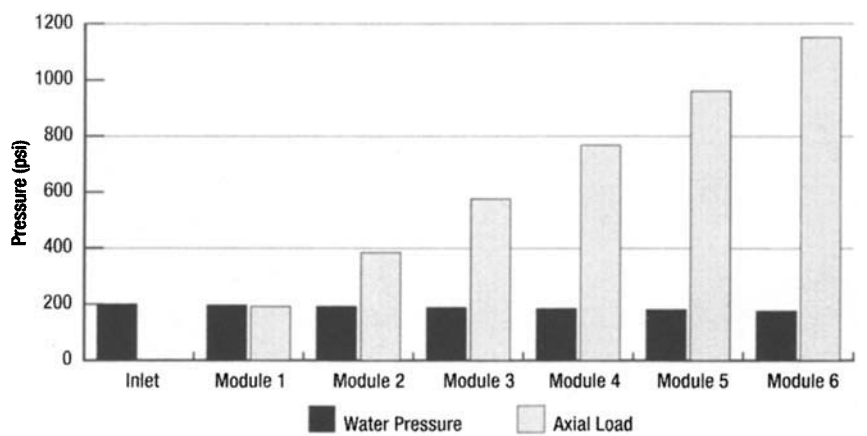
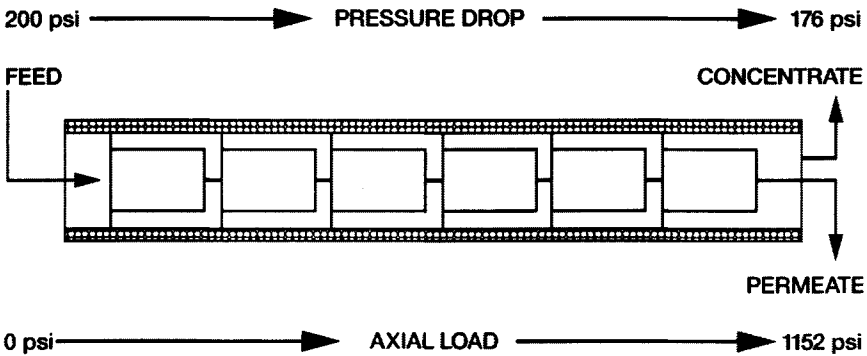


Figure 11.5 Axial pressure load on 8-inch diameter membrane modules operating at 200 psig. Assumes six, 8-inch modules per pressure vessel.

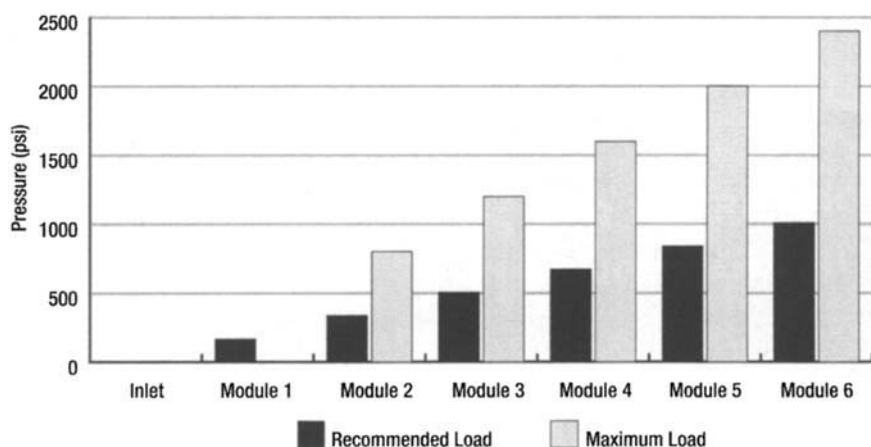


Figure 11.6 Comparison of axial pressure load on end module with recommend 3–4 psig pressure drop per module and a maximum recommended pressure drop of 8.3 psig per membrane module. Assume six, 8-inch diameter modules per pressure vessel.

physical damage to the membrane itself (see Chapter 14.8.1). Figure 11.5 shows how the pressure drop on an 8-inch (20 cm) diameter membrane module operating at 200 psig increases the axial load on the modules themselves. Assuming a “good” pressure drop of 4 psig per membrane module, the axial pressure on the end of the last module is over 1,150 psig. Figure 11.6 shows the different axial loads for 8-inch diameter modules operating at 200 psig with a recommend pressure drop of 3 – 4 psig per membrane module (21 psig drop per single stage of 6 modules in series), and a maximum recommended pressure drop of 8.3 psig per membrane module (50 psig maximum drop per single stage of 6 modules in series). At 50 psig pressure drop, the axial load on the end of the last module is 2400 psig.

As a result, pressure drop should also be considered in making the determination when to clean the membranes to avoid physical damage to the modules. Membranes should be cleaned when the pressure drop increases by 10% to 15% from initial operating conditions.

Equation 11.6 can be used to normalized the system pressure differential.⁵

$$\frac{\text{Normalized Differential Pressure}}{\text{Pressure}} = \frac{\Delta P_{\text{actual}} \times (2 \times CF_{\text{su}} + PF_{\text{su}})^{1.5}}{(2 \times CF_{\text{actual}} + PF_{\text{actual}})^{1.5}} \quad (11.6)$$

where

ΔP_{actual} = actual differential pressure

CF = concentrate flow

PF = permeate flow

"su" = subscript for start up

Normalized pressure drop and NPF should be monitored simultaneously, and the membrane cleaned when the first of these measures reaches the 10% change in performance as compared to initial operating performance.

11.3.2 Normalization Software

Normalization calculations can be tedious, so membrane manufacturers have made available, at no charge, normalization software for their specific membranes. Some chemical vendors and other membrane consultants have software as well, which may or may not be available for public use. Some even have hardware that will download the appropriate data from the PLC or discrete probes to eliminate the need for manual entry of the data. Operators will be able to use the normalization software to review the normalized data, schedule cleanings, and optimize the RO operation.

Table 11.2 lists the typical data inputs required by the normalization software programs. These inputs cover a multi-stage RO skid as if it were a single stage in terms of calculating normalized data. Some programs have the ability to handle interstage data, which allows the user to analyze each stage separately (typically limited to 2 stages). The ability to analyze individual stages aids the user in determining whether fouling is occurring in the first stage or scaling is occurring in the second stage.

Virtually all normalization programs will calculate the normalized permeate flow, normalized salt rejection and/or passage, and differential pressure (some programs normalize it, some do not). Some programs also include net driving pressure as an output as well as the following outputs:

- Recovery
- Concentration factor
- Salt passage
- Salt rejection
- Average system flux
- Average feed osmotic pressure

Table 11.2 Typical input data required by normalization software programs.

Input Data	Raw water	Feed	Permeate	Reject
Date and time		X		
Flow		X	X	X
Pressure			X	X
Conductivity		X	X	
Turbidity	X	X		
ORP		X		
SDI	X	X		
Selected events*		X		

*Cleaning, membrane replacement, etc.

Hydranautics' ROData program includes the normalized water transport coefficient and the normalized salt transport coefficient. The water transport coefficient corresponds to the permeability coefficient, A , in equation 4.1. The salt transport coefficient corresponds to the permeability coefficient, K , in equation 4.2. The coefficients should remain constant over normal (ambient) operating conditions. Changes in the coefficients are indicative of changes on or to the membranes. An increase in the water transport coefficient generally implies that the membranes are degraded, while a decrease in the coefficient means the membranes are fouling or scaling. Similarly, an increase in salt transport coefficient means the membranes are degraded or scaled while a decrease typically means that the membranes are fouled.

The normalization programs provide graphs for the performance variables. Some programs offer the ability to combine various data curves on the same graph. For example, Toray Trak by Toray Membrane USA has the ability to include four (4) curves on the same graph. Figure 11.7 shows normalized permeate flow and normalized salt rejection as functions of time using Toray Trak. Figure 11.8 shows the normalized differential pressure as a function of time. Multiple curves, whether on the same graph or not, allow the user to compare different performance parameters and aid in troubleshooting the system, as trends are easier to discern. For example, the loss in permeate flow shown in Figure 11.5,

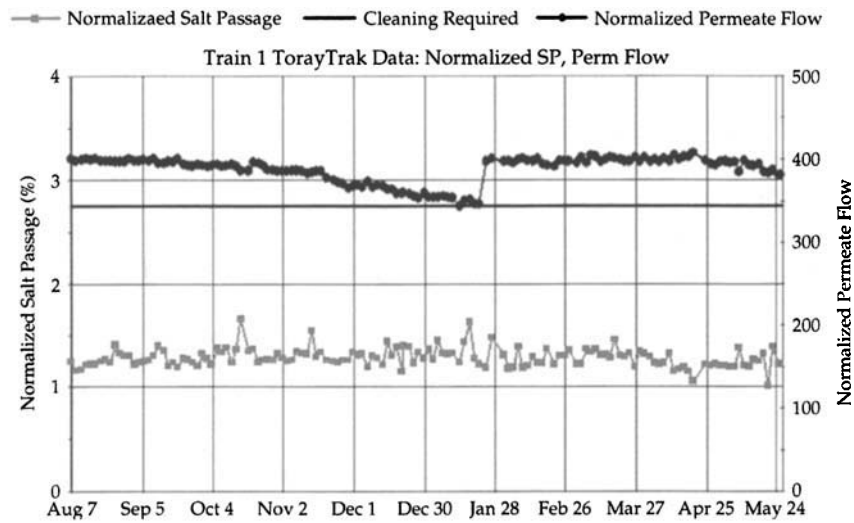


Figure 11.7 Normalized permeate flow and normalized salt rejection as functions of time. *Data courtesy of Madalyn Epple, Toray Membrane, USA, Inc.*

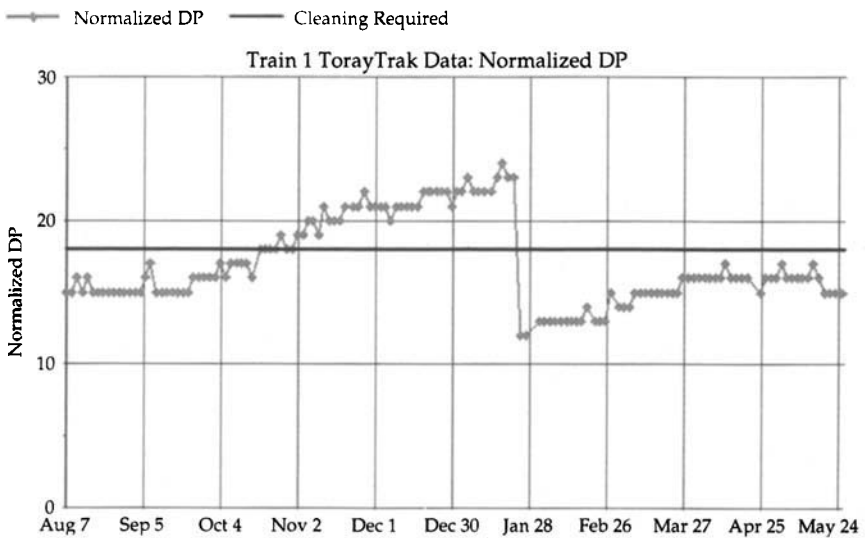


Figure 11.8 Normalized differential pressure as a function of time. *Data courtesy of Madalyn Epple, Toray Membrane, USA, Inc.*

coupled with the corresponding increase in differential pressure, Figure 11.6 and the constant salt rejection, Figure 11.5, could be caused by membrane fouling rather than scaling (see Chapter 13.3).

Both graphs (Figure 11.7 and 11.8) show the suggested time to clean the membranes based on initial performance. The increase in normalized differential pressure reached the “cleaning required” condition after about 3 months of operation, which is typical for a system with good pretreatment. However, this cleaning indicator was overlooked in favor of the normalized permeate flow, which did not drop to its cleaning indicator condition until about five and one half months after start up. Fouling does not appear to have been permanent, however, as both the normalized permeate flow and differential pressure returned to start up performance following cleaning.

Other outputs available from at least some normalization programs include:

- System recovery: this can be compared to the hand-calculated recovery as well as the recovery based on rejection of species such as sulfate (which is rejected to a high degree);
- Net pressure driving force: looks at the effect of concentration on the pressure driving force for water flux;
- Permeate and feed/concentrate average osmotic pressures: used to calculate the net pressure driving force;
- Specific flux: calculated by dividing the flux by the net driving pressure. The higher the specific flux, the greater the permeability of the membrane. Low permeability corresponds to fouling, scaling, or compaction.
- Event identification: Allows the user to input events such as cleaning, new membranes, etc., and the performance graphs will indicate that an event has occurred by placing a marker on the graph corresponding to the date the event occurred. This helps in identifying any step changes that may occur in performance after an event.

11.4 Preventive Maintenance

Preventative maintenance (PM) is critical to RO operations. It may require additional staff time and capital, just as the pretreatment to RO requires additional capital, but in the long run, PM will pay for itself in improved RO operations, including longer membrane life. Successful RO operation requires diligent PM, so PM tasks must

be assigned to be completed on a regular basis. Table 11.3 lists a suggested PM schedule for general maintenance tasks.

Tasks listed in the Table 11.3 are critical for the performance of the system and to keep total operating costs, including membrane cleanings and replacements, to a minimum.

Table 11.3 General preventative maintenance schedule (adapted from Anne Arza, Nalco company).

Preventive Maintenance Item	Daily	Weekly	Monthly	Quarterly
Daily Logs	X			
Normalization	X	X*		
SDI	X	X*		
Water Analysis (feed permeate, reject)			X	
Mass Balance			X	
Check 5-micron Prefilters		X**		
Check and Calibrate Critical Sensors			X	
Check and Calibrate all Sensors				X
Check Pretreatment Unit Operations (e.g., filters, softeners, etc)			X	
Calibrate Chemical Feed Pumps				X
Pump Maintenance (all)			X	
Elution Studies (if necessary)				X+
Membrane Cleaning				X++

* after start-up and system equilibrium

** Biweekly

+ Semi-annually

++ Or as required per normalized data

- *Daily logs:* Collecting data is perhaps the most fundamental aspect of PM. Without data, there is no way to determine the conditions of the membranes. See Chapter 11.2 for details about the type of data to collect.
- *Normalization:* Should be conducted on a daily basis following start-up or the installation of new membranes. Once the system has reached equilibrium, which could take up to 4 to 6 weeks, normalization can be done on a weekly schedule. See Chapter 11.3.1 for more details about data normalization.
- *SDI:* Silt density index tests should be conducted on a daily basis following start up, but once the system has reached equilibrium, weekly readings can be taken. Note that if the make-up water is a surface source, SDI may need to be taken daily to catch changes due to weather conditions (e.g., heavy rains), ship traffic, and so forth.
- *Water analyses:* See Chapter 7 for details on what species to analyze.
- *Mass balance:* Mass balance is an effective way to troubleshoot an RO system to determine whether species are being deposited on the membranes. See Chapter 14.4 for information about mass balances.
- *Check cartridge filters:* The cartridge prefilters should be checked biweekly for pressure drop across them and to be sure they are seated properly. High pressure drop means it is time to replace the filters. Improper seating of the filters will lead to particulates bypassing the filters and fouling or abrasion (and destruction) of the membranes.
- *Clean and calibrate critical sensors:* Critical instruments include the ORP and pH sensors used to make sure the RO feed water is acceptable.
- *Clean and calibrate all sensors:* Instruments must be calibrated on a regular basis. Improper sensor readings will lead to inaccurate normalization and present a false picture as to how the RO system is functioning.
- *Calibrate chemical feed pumps:* Chemical feed systems should be calibrated on a regular basis to make sure the required dosage of chemical is being fed.

- *Check pretreatment unit operations:* The integrity of filters, softeners, tanks, and other pretreatment systems should be checked. Performance of each should also be evaluated and modifications made as required.
- *Elusion studies:* Elusion studies should be conducted on softeners on a quarterly basis or more frequently, if required.
- *Membrane cleaning:* RO membranes should be cleaned based on normalized performance. Cleanings should not be more frequent than every three months if the pretreatment system is adequate. RO systems with more frequent cleanings could benefit from upgrades to the pretreatment system.

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Performance Degradation

Performance of an RO system, specifically the permeate flow rate, salt rejection and pressure drop, is a function of membrane fouling, scaling, and degradation, as previously discussed (See Chapter 11.3). This chapter covers the detailed effects of membrane fouling, scaling, and degradation have on normalized product flow, normalized salt rejection, and pressure drop.

12.1 Normalized Permeate Flow

Normalized permeate flow (NPF) is a function of the average applied transmembrane pressure, the osmotic pressures of the feed and permeate, and temperature, as shown in Equation 11.1. Factors that cause an increase or decrease in the NPF are discussed below.

12.1.1 Loss of Normalized Permeate Flow

Membrane fouling and scaling can both lead to a loss in normalized permeate flow. Additionally, membrane compaction will result in decreased permeate flow as well.

12.1.1.1 *Membrane Fouling*

Membrane fouling involves the deposition of suspended solids, including bacteria, on the membrane or components within the membrane module. These foulants form a layer on the surface of the membrane that becomes an additional barrier for water to flow through to the permeate side of the membrane. Hence, if the feed pressure is held constant, the permeate flow will decrease.

In general, operators focus on observed permeate flow and adjust operating variables to deliver the required permeate flow. Thus, if the permeate flow is decreasing due to fouling, the operating pressure is usually increased to overcome the additional barrier to transport and to maintain a constant observed permeate flow.

As discussed in Chapter 11.3.1.1, membrane fouling will result in a negative slope for the NPF (see Figure 11.1).

12.1.1.2 Membrane Scaling

Membrane scaling involves the deposition of saturated salts on the surface of the membrane, typically in the later stages of the RO system. Scale forms a layer on the surface of the membrane that becomes an additional barrier for water to have to flow through to the permeate side of the membrane, similar to that described above for foulants.

Typically, operating pressure is increased to adjust for the loss in observed permeate flow due to scaling, so observed permeate flow is not a good indicator of membrane scaling. However, the normalized permeate flow will reflect the need to increase pressure with scaling, and register this as a negative slope in the normalization curve. Thus, membrane scaling will result in a decrease in NPF.

12.1.1.3 Membrane Compaction

Membrane compaction involves the compressing of the membrane itself that, in essence, makes the membrane “denser” or thicker and reduces the flow and salt passage through it. Compaction can occur under higher feed pressure, high temperature, and water hammer. (Water hammer occurs when the RO high-pressure feed pump is started and there is air trapped in the membrane modules—see Chapter 6.2.) Most brackish water membranes, when operated properly, experience a minimal degree of compaction. However, seawater membranes and cellulose acetate membranes at pressures greater than about 500 psig, can experience significant compaction.¹ Compaction will cause a decrease in NPF.

12.1.2 Increase in Normalized Permeate Flow

An increase in normalized permeate flow is typically the result of a leak, either due to a breach in the membrane itself or because of problems with the membrane module hardware, or to exposure to oxidizers such as chlorine.

12.1.2.1 Membrane Degradation

Membrane degradation involves the loss of membrane polymer integrity. This can be a result of membrane oxidation, where the oxidizer attacks the membrane polymer, leading to aromatic

ring substitution and chain cleavage.¹ In this case, feed water is allowed to pass into the permeate leading to an increase in permeate flow and a decrease in product quality. Oxidation degradation typically occurs at the lead membranes, as these membranes are exposed to the oxidizer first, and, in many cases, the oxidizer has been reduced before it can reach the later membranes. Typical oxidizers include free chlorine (and other halogens), chlorine dioxide, ozone, hydrogen peroxide, and so on (see Chapters 8.2.1 and 8.2.2).

Membranes can also be oxidized in the presence of iron, manganese, and other metals. These metals catalyze the oxidation of RO membranes. This type of oxidation tends to involve the entire RO skid rather than focus on the lead membranes. Again, when this type of degradation occurs, feed water passes into the permeate, resulting in an increase in permeate flow and a decrease in product quality.

Exposure to high temperature at pH extremes can hydrolyze the membranes, leading to loss of membrane integrity. (See Chapter 4.2.1, Table 4.2, and Chapters 9.2, 9.8, and 13.2 for more detailed discussions on the effect of temperature and pH on polyamide composite membranes.) Hydrolysis also tends to involve the entire RO skid rather than focus on only the lead membranes. Just as with oxidation of the membrane, feed water will pass into the permeate resulting in an increase in permeate flow and decrease in the product quality.

Membrane degradation can also be a physical phenomenon. Particles, such as granular activated carbon fines, abrade the membrane surface and cause microscopic tears in the membrane materials through which feed water can breach the membrane, and increase the permeate flow. Also, excessive permeate back-pressure can lead to ruptures in the glue lines holding the membrane leaves together, again leading to breaches through which feed water can enter the permeate and the permeate flow increases (see Figure 12.1).

12.1.2.2 *Hardware Issues*

Breaches in hardware will allow feed water pass into the permeate. Common problems include O-ring leaks and leaking module product tubes, both of which can be damaged during installation. Product tubes may also be damaged under conditions of high pressure drop due to excessive fouling or scaling. Water hammer may cause damage to the permeate tube as well.

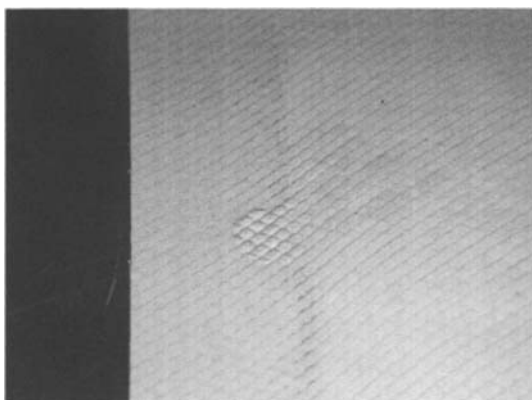


Figure 12.1 Glue line breaches where feed water and permeate can mix.

12.2 Normalized Salt Rejection

Normalized salt rejection is a function of the concentration driving force across the membrane, as shown in Equation 11.3. Factors that lead to loss or increase in salt rejection are discussed below.

12.2.1 Loss of Salt Rejection

Membrane scaling and degradation can lead to a loss in normalized salt rejection as can breaches O-rings and permeate tube.

12.2.1.1 *Membrane Scaling*

Membrane scaling involves the deposition of saturated salt(s) onto the surface of the membrane or components within the membrane module. This scale forms a layer on the surface of the membrane where the concentration of the saturated salt(s) is higher than the concentration of the salt(s) in the bulk solution. The concentration that the membrane is exposed to is thus higher than that recorded for the bulk solution. Since the actual percent rejection exhibited by the membrane remains fairly constant during nominal, industrial-type applications, the salt passage also remains the same. However, since the concentration of salt(s) is higher at the membrane surface, the actual amount of salt(s) that passes through the membrane is higher than would be expected based on the bulk concentration. Hence, the apparent salt passage

increases and the apparent rejection decreases. This phenomenon is registered as a loss in normalized salt rejection.

12.2.1.2 Membrane degradation

Membrane degradation involves the loss of membrane integrity (see Chapter 12.1.2.1). Because of the loss in integrity, feed water is allowed to pass into the permeate leading to an increase in permeate concentration. Thus, salt passage increases while the salt rejection decreases.

12.2.1.3 Hardware Issues

Breaches in hardware can allow feed water to mingle with permeate. As is the case with an increase in water flux, breaches in O-rings and permeate tubes will allow high concentration feed water to mingle with low concentration permeate, thereby increasing the concentration of the permeate (see Chapter 12.1.2.2). The overall salt rejection will decrease and salt passage will increase.

12.2.2 Increase in Salt Rejection

Increases in salt rejection are typically due to membrane compaction (see Chapter 12.1.1.4). As the membrane becomes denser due to compaction, the passage of salts through the membrane is reduced, leading to a loss in salt passage and an increase in salt rejection.

Note that brief, initial exposure to chlorine under tightly controlled conditions can also lead to an increase in salt rejection for some membranes (while longer exposure leads to a decrease in salt rejection).³ The reasons for this are not clear, but could involve changes to the membrane surface charge which results in a repelling of anionic compounds, decreasing their passage and increasing the rejection.³

12.3 Pressure Drop

Pressure drop measures the loss in pressure from the feed to the concentrate. In effect, it measures the loss in driving force for water across the membrane (see Chapter 12.3.1.3 and Equation 12.6). Factors that result in an increase or decrease in pressure drop are discussed below.

12.3.1 Loss in Pressure Drop

It is unusual to find a loss in pressure drop under nominal operating conditions for most industrial applications.⁴ If a loss in pressure drop is recorded, it is typically a result of faulty instrumentation.

12.3.2 Increase in Pressure Drop

A number of factors can lead to high pressure drop, including membrane scaling, colloidal fouling, and microbial fouling. These three factors all involve deposition of material onto the surface of the membrane as well as onto components of the membrane module, such as the feed channel spacer. This causes a disruption in the flow pattern through the membrane module, which, in turn, leads to frictional pressure losses or an increase in pressure drop.

High pressure drop causes disruptions to the system hydraulics. Because of the high pressure drop, the lead membranes tend to operate at very high fluxes while the lag membranes operate at low flux. This increases the rate of membrane fouling for both the lead and lag membranes. Lead membranes foul faster because more water is forced to the membrane module faster and the rate of contaminant accumulation in the boundary layer on the membrane surface increases. The lag membranes, on the other hand, experience low flows since most of the water is removed through

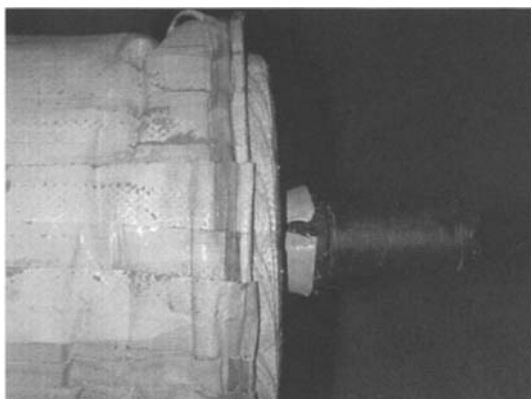


Figure 12.2 Damaged membrane module due to high pressure drop.

the lead membranes. In this case, the thickness of the boundary layer increases, leading to faster rates of fouling.

High pressure drop can also lead to damage of the membranes and membrane modules themselves. The loss in feed water pressure is translated into pressure down the axis of the membrane module. When the pressure drop gets very high, greater than about 50 psig per 6-element stage, the membrane modules can telescope which can physically crack and tear the membrane or compact the fiberglass module shell, as shown in Figure 12.2 (see Chapter 3.6, Figure 3.5, Chapter 11.3.1.3, and Figure 11.6).

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Off-Line Operations

Operations that are conducted while the RO system is off line are just as important as on-line operations in keeping an RO system functioning well. Off-line operations covered in this chapter include system flush, membrane cleaning, and membrane lay-up.

13.1 System Flush

System flushes are typically used when an RO system goes off-line, comes back on-line, and during stand-by mode. The purpose of the off-line and stand-by flushes is to rid the feed/concentrate side of the membrane of either high concentrations of feed water species or to stir up materials that may have settled on the membrane during down time. The on-line flush (when the membranes come back on line) is to reduce the conductivity in the RO permeate before sending the permeate on to further processing or to the ultimate use. Flush water is typically sent to drain.

It is important during flush to keep the permeate line wide open to prevent delamination of membranes at the concentrate end of the pressure vessel (see Figure 13.1). Delamination occurs when the permeate pressure is higher than the pressure of the feed side of the membrane. During fouling conditions, the feed-side pressure drop can be significantly higher than the 21 – 24 psig for a 6-module pressure vessel under “clean” conditions. If the permeate line is closed, the potential exists for the permeate pressure to exceed the feed pressure due to the pressure drop on the feed-side of the membrane. If this difference is greater than about 10 – 15 psi, delamination can occur.¹

13.1.1 Off-Line Flush

The off-line flush is conducted when the membranes are brought off line for any reason. The flush is used to displace the high concentration of solids in the feed/concentrate side of the membrane with lower-concentration water (feed water under minimal pressure so as

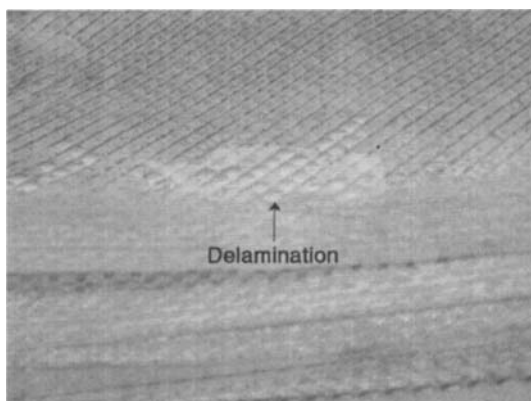


Figure 13.1 Membrane delamination caused by high permeate pressure.

not to concentrate the flush water). A flush of 3 to 5 minutes is usually sufficient. Any permeate generated is sent to drain. Typically, pretreated (softened) feed water is used for the flush, although permeate water is preferred (as a last resort, low-pressure raw feed water can be used). The flush water should be free of pretreatment chemicals, so any chemical injection systems should be stopped before flushing. After the flush is completed, all feed valves should be completely closed. If the concentrate line drains to a level that is below that of the pressure vessels, the vessels might be emptied by a siphoning effect. To prevent this, an air break should be installed in the concentrate line at a position higher than the highest pressure vessel.

The flush can be programmed into the PLC and thus occurs automatically when the RO skid shuts down or it can be manually initiated. Note that not all RO systems are equipped with this flushing feature. Without this feature, fouling and scaling of membranes will be exacerbated, particularly if the RO unit cycles on and off on a regular basis.

13.1.2 Return to Service Flush

The on-line flush takes place when the RO skid is returned to service. The objective of the flush is to remove particulates and salts that have settled on the membrane surface while the skid was off line as well as to bring down the concentration of the RO permeate. Figure 13.2 shows the concentration of the RO permeate as a function of flush

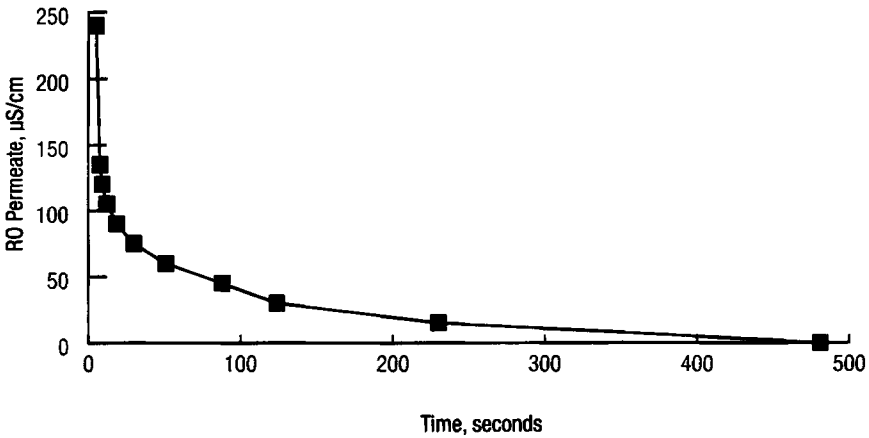


Figure 13.2 Permeate concentration as a function of flush time after membrane stand-by (250 ppm micro Siemens per centimeter feed conductivity). *Courtesy of Jonathan Wood and UltraPure Water® Journal, www.ultrapurewater.com, March, 2009.*

time after stand-by (assumes a feed water conductivity of 250 micro Siemens per centimeter).² Typically, pretreated feed water is used for the flush; there is no need to use higher-quality water for this flush. Pressure is generally at nominal operating condition, and any permeate that is generated is sent to drain.

The on-line flush can also be programmed into the PLC and occurs automatically when the RO skid is brought back on line. Most RO systems are equipped with this feature.

13.1.3 Stand-by Flush

The stand-by flush is used intermittently when the RO skid is off line in stand-by mode. It can also be used during extended lay-up of the skid. The objective is to remove particles and salts that have collected on the membrane surface while the membranes are idle. This minimizes the potential for membrane fouling and scaling while the membranes are at rest.

The frequency for this type of flush is application and environment dependent. Flushing may be programmed to occur every 15 minutes to once every 2 hours, depending on how concentrated the pretreated feed water is and how high the temperature is. Higher feed water concentrations and higher temperatures require shorter

intervals between flushes. Flushes of 3 to 5 minutes are sufficient. Permeate water is preferred to flush the membranes, although softened feed can be used (raw feed water can be used as a last resort). Pressure is minimal to avoid concentrating the water used to flush the membranes. Any permeate generated during the flush cycle is sent to drain.

13.2 Membrane Cleaning

Membrane cleaning is one of the most important aspects of membrane operations. Regardless of how good the pretreatment and hydraulic design of an RO system, membranes will eventually foul and/or scale. Timely and effective cleaning is necessary to keep membranes free of foulants and scale, which can result in longer cleaning intervals, and longer membrane life, both of which save money, time, and the environment.

13.2.1 When to Clean

Membranes should be cleaned when the normalized permeate flow drops by 10% – 15% from initial stabilized performance (see Chapter 11.3.1.1), or when the differential pressure increases by 10% – 15% (see Chapter 11.3.1.3). Ideally, a cleaning is scheduled when the performance changes by 10% and should be completed by the time the performance has changed by 15%. Waiting too long to clean can result in irreversible fouling and/or scaling of the membrane. Membranes with good pretreatment can expect to clean about 4 times per year or less.

Membranes are typically not cleaned due to a drop in salt rejection. This is because in most instances, there is a mechanical explanation for the drop in salt rejection (see Chapter 12.2.2). However, in the case where membrane scaling is responsible for a drop in salt rejection, normalized permeate flow is generally the first indicator of this phenomenon (see Chapter 11.3.1.2)

Figure 13.3 compares the projected performance of an RO membrane that has been cleaned on time (performance decline within 15% of initial performance) with that for membranes that were cleaned after performance had dropped more than 15%. As the figure shows, cleaning on time results in better cleaning efficacy and in longer intervals between cleanings. Both of these outcomes result

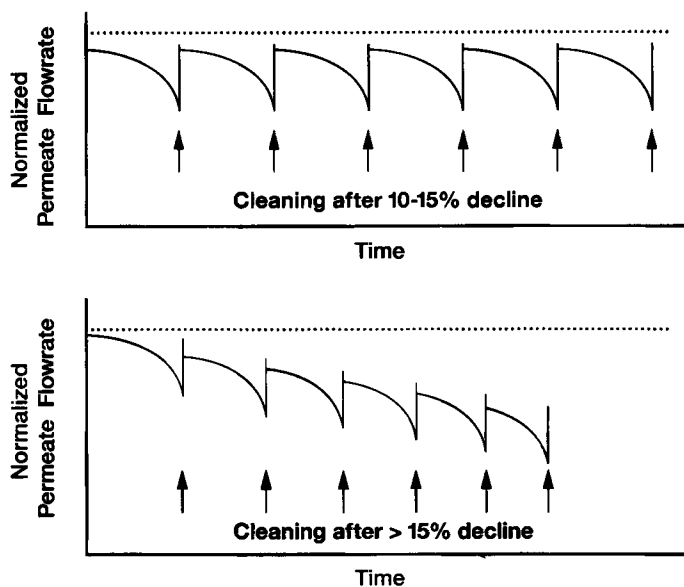


Figure 13.3 Comparison of cleaning when performed within 15% of performance decline and when performance has fallen more than 15% from start-up conditions.

in lower operating and maintenance costs for the system. Waiting too long to clean can cause some of the foulants and/or scale to become permanently attached to the membranes or module components. Once foulants or scale have deposited onto the membrane or module components, they attract more materials, exacerbating the problem, and leading to a more rapid decline in performance and, therefore, more frequent cleaning episodes.

13.2.2 How to Clean

Membrane manufacturers and cleaning-chemical vendors typically have cleaning procedures formulated for their specific product(s). While it is impossible to review all cleaning procedures here, a basic cleaning procedure is presented which can be modified to suit the membrane, the cleaning chemical(s), and the specific contaminant(s) to be removed:

1. *Prepare the cleaning solution.* If the solution is in liquid form, it just needs to be pH-adjusted and heated. If the solution is in dry form, it needs to be mixed (using

RO-permeate or better quality water), pH-adjusted and heated

2. *Introduce the solution to the RO skid.* It is important to clean each stage individually, so that foulants formed in the first stage do not carry over to the second stage and scale formed in the second stage does not carry over to the first stage. The solution should be introduced at a flow rate of about 3 to 5 gpm per pressure vessel.³ For an 8-pressure vessel stage, the total introduction rate would be no more than about 40 gpm. This slow flow rate gives the user enough time to view the cleaning solution and observe changes in color. If the cleaning solution changes color (e.g., becomes dark), it should be discarded and a fresh batch should be prepared.
3. *Recirculate the solution.* The solution should be recirculated at about 35 gpm per pressure vessel. In other words, for an 8-pressure vessel stage with 8-inch diameter membrane modules, the recirculation rate should be 8 times 35 gpm or 280 gpm. Recirculation should be conducted using as little pressure as possible, thereby minimizing the formation of permeate. If permeate is generated, it increases the likelihood of re-deposition of removed species on the membrane.¹ If the cleaning solution comes out dark, it should be discarded and a new batch should be prepared. Temperature and pH should be monitored and adjusted during the recirculation as needed. Recirculation should continue for about 45 to 60 minutes.
4. *Soak the membranes.* Membrane should be allowed to soak in order to give the cleaning solution time to "loosen" material on the membrane surface and to penetrate biofilm. In some cases, a soak of one hour is sufficient. In other cases an overnight soak may be required. During longer soak periods, use the cleaning skid pump to recirculate the solution at about 10% of the typical recirculation rate to help maintain temperature. Note that longer periods of soaking is not a substitute for using the temperature and pH as recommended by the manufacturer for efficient cleaning.
5. *High-flow recirculation.* The solution should be recirculated at about 40 – 45 gpm per 8-inch pressure vessel

for about 45 minutes. Note that higher flow rate recirculation may result in pressure drop issues. The solution should be recirculated at no more than about 50 psig for 8-inch diameter, six-element pressure vessels.

6. *Flush the membranes.* Membranes should be flushed following cleaning using RO-permeate quality or better water. Pretreated feed water should not be used as components may interact with the cleaning solution and precipitation of foulants may occur in the membrane modules. The minimum flush temperature should be 20°C.¹
7. *Repeat steps 1 through 5* using the next cleaning chemical(s).

Factors that effect cleaning efficacy include:

- *Chemicals used:* using appropriate cleaning formulations will greatly enhance the efficacy of the cleaning procedure. The concentration of cleaning chemicals is also important to the overall effectiveness of cleaning.⁴
- *Temperature:* cleanings conducted at higher temperatures are more effective at removing materials from membranes.
- *pH:* extremes in pH (both high and low) are more effective in cleaning membranes. However, temperature and pH limits for cleaning should be followed to prevent damaging the membranes or module components (see Table 13.1).⁴
- *Recirculation rate:* the velocity of recirculation will impact the ability of the cleaning solution to remove debris from the membrane. Higher flow rates (see step 5 above) are usually necessary to scour debris off of the surface of a membrane.
- *Time:* the time spent cleaning and soaking the membrane is vital to the overall success of cleaning. Sufficient time must be devoted to the cleaning process to ensure that the cleaning solution has time to penetrate foulants and scale and that the recirculation flows scour as much of the debris off of the membranes as possible.

These factors should be monitored and recorded at 15 minute intervals during the entire cleaning process. Temperature, pH, and recirculation rate should be adjusted to maintain the initial cleaning specification which will ensure the best cleaning action.

13.2.3 Cleaning Chemicals

Effective cleaning is a function of pH, temperature, and cleaning solution(s). Cleaning is most effective when it is focused on the specific fouling or scaling problem experienced by the membrane. If the “wrong” cleaning chemicals are used, the situation can become worse.¹ If in doubt, most membrane manufacturers recommend conducting the alkaline cleaning first.¹ Acid cleaning should only be used first if it is known that calcium carbonate or iron oxides are present in the membrane modules. If acid cleaning were to be used first, and microbial or organic fouling was present, the fouling situation may become irreversible.

Cleaning solutions are usually classified by pH. There are high-pH, low-pH, and neutral-pH cleaners. Increasing temperature and pH extremes together enhance the effectiveness of cleaning. However, there are limits to temperature as a function of pH, as shown in Table 13.1 for Dow Water and Process Solutions polyamide composite membranes (membrane manufacturers should be consulted prior to cleaning to confirm their temperature/pH limitations).¹ At high temperature and pH extremes, a conversion of the membrane takes place, resulting in a loss of performance. The mechanism for this change in the membranes is not clearly understood.⁵

Table 13.1 Temperature and pH limitations for dow water solutions polyamide composite membranes.¹

Temperature (°C)*	pH
25	1 – 13
35	1 – 12
45	1 – 10.5

For all brackish water membranes (BW30, BW30LE, LE, and XLE)

* for temperatures greater than 50°C, Dow-FilmTec recommends that they be contacted

In addition to acid or caustic, cleaners will sometimes have other compounds that are effective in removing and suspending materials off of the membrane. These compounds include metal chelating agents, surfactants, and enzymes. These compounds clean a membrane by one of three mechanisms:⁴

- physically removing the foulants or scale off of the surface of the membrane
- changing the morphology of the foulant or scale such that further accumulation is discouraged
- changing the surface chemistry of the fouling layer to, again, dissuade additional accumulation.

The user can mix these cleaning formulations or some may be available pre-packaged from a vendor. *For vendor-provided pre-packaged cleaners, the user should determine the compatibility of the cleaning formulation with the specific membranes being cleaned to avoid potential fouling or degradation of the membrane with the components in the cleaning solution.* Some specific cleaning formulations are described below.

13.2.3.1 High-pH cleaners

High-pH cleaners are typically used to address removal of the following species:

- Biofilm fouling
- Organic fouling
- Silica scale
- Colloidal material fouling
- Sulfate scale

High-pH cleaners are generally formulated with a surfactant (detergent) such as sodium dodecylsulfate (Na-DSS) or sodium laurel sulfate. *Note that cationic surfactants should not be used, as irreversible fouling of the membrane may occur.* The surfactant serves to lift solids of the membrane and support them in the solution. (Antifoams may be added to suppress the foaming action of the surfactant.) Some high-pH cleaners may also be formulated using a chelating agent such as sodium EDTA (ethylenediaminetetraacetic acid). The chelating agent serves to maintain salts in solution. Sequestering agents may also be added to bind up calcium and other scale-forming

Table 13.2 Sample high-pH cleaning formulations for polyamide, composite membranes.

	Sulfate scale	Organic Fouling	Biofouling	Silica
0.1 % caustic, pH 12, 35°C	X	X	X	X
1.0 % Sodium EDTA*, pH 12, 35°C	X			
0.025% sodium DSS**, pH 12, 35°C		X	X	X
2.0% sodium TPP+ and 0.8% sodium EDTA, pH 10.0, 45°C	X	X		
2.0% sodium TPP+ and 0.25% sodium DDBS++, pH 10.0, 45°C		X		

* ethylene diamine tetraacetic acid

** dodecylsulfate

+ Tripolyphosphate

++ dodecylbenzene sulfonate

cations. Table 13.2 lists four sample, high-pH cleaning solutions formulations applicable to polyamide, composite membranes.^{1,6}

Figure 13.4 shows the effect of high pH on the removal of biofilm.¹ As the figure shows, pH of 12 is much more effective at restoring permeate flow than lower pH.

Note that performance for the RO unit may be different after a high pH cleaning. Specifically, the flux may be higher and the rejection may be lower. This is a result of the membrane “loosening” at high pH and is a temporary condition. Performance should return to nominal within a couple of days.

13.2.3.2 Neutral-pH Cleaners

Neutral-pH cleaners are usually used to address microbes. Most of these cleaners use a non-oxidizing biocide such as DBNPA (dibromonitripropionamide) or isothiasolin, and are pre-packaged by vendors.

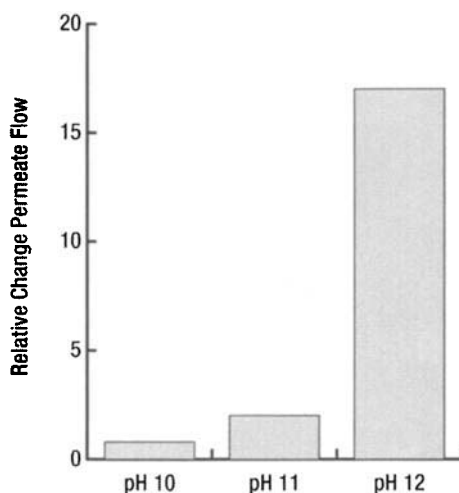


Figure 13.4 Effect of pH on ability of cleaning solution to remove biofilm and restore membrane performance. *Courtesy of Dow Water and Process Solutions.*

13.2.3.3 Low-pH Cleaners

Low-pH cleaners are typically used to address calcium carbonate scale and iron oxide deposition. These cleaners are usually formulated using only acid, such as acetic, hydrochloric, or sulfamic. Figure 13.5 shows the effects of temperature and pH on the removal of calcium carbonate from a membrane.⁷ As the figure shows, lower pH, and higher temperatures are more effective at restoring permeate flow than higher pH and lower temperatures.

Hydrogen peroxide or a combination of hydrogen peroxide and peracetic acid can be used to address biologically fouled RO membranes when operated at lower pH. The following precautions should be taken when using hydrogen peroxide:

- Dilute commercial products to 0.2wt% with RO permeate. Higher concentration will result in membrane damage (higher salt passage).
- Maintain temperature less than 25°C. Higher temperature will result in membrane damage.
- Remove all transition metals from the solution. The presence of metals will catalyze the degradation of the membrane.

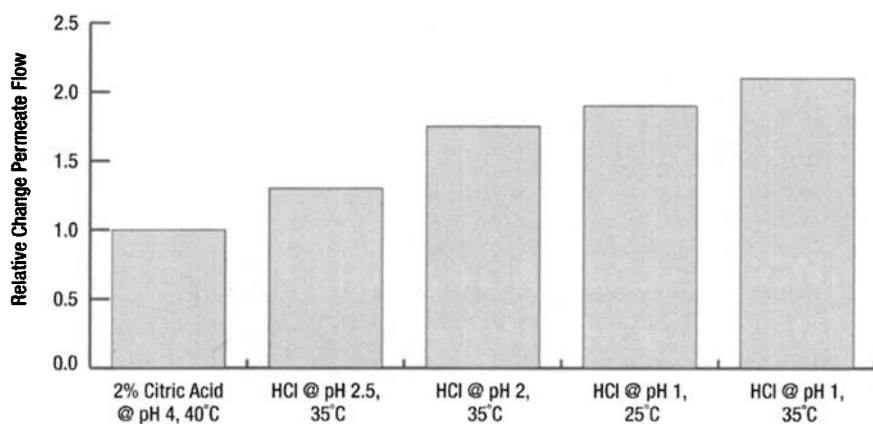


Figure 13.5 Effects of temperature and pH on removal of calcium carbonate from an RO membrane. *Courtesy of Dow Water and Process Solutions.*

- Recirculate diluted hydrogen peroxide solution for 20 minutes.
- Maintain a pH of 3-4 for optimal efficacy and longest membrane life.
- Completely clean the membranes with high pH and low pH cleaners prior to recirculating hydrogen peroxide.

As with high pH cleaning, the performance of the RO membranes after cleaning may temporarily change from nominal. In the case of low pH, the membrane essentially becomes “tighter,” exhibiting lower flux and higher rejection for up to a few days after the cleaning is completed.

13.2.3.4 Cleaners for Specific Foulants and Scale

For stubborn foulants and scale, there are some preferred cleaning solutions that may improve the efficacy of cleaning. Some of these generic solutions are listed in Table 13.3.^{1,8} Check with specific membrane manufacturer for limits on pH and temperature.

13.2.4 Cleaning Equipment

Reverse osmosis cleaning equipment is simple and straightforward. Figure 13.6 shows a schematic of a cleaning skid. The skid can be

Table 13.3 Preferred cleaning solutions for specific foulants and scale.^{3,8}

Species	Dow-FilmTec	Hydranautics*
Sulfate Scale	0.1% caustic, pH 12, 30°C	2.0% sodium tripolyphosphate, pH 10, 45°C
Carbonate Scale	0.2% hydrochloric acid, pH 2, 30°C	0.5% hydrochloric acid, pH 2.5, 45°C
Silica Scaling	0.1% caustic, pH 12, 35°C	0.1% caustic, pH 11.5, 35°C
Iron Fouling	1.0% sodium hydrosulfate, pH 5, 30°C	1.0% sodium hydrosulfate, pH 11.5, 35°C
Organic Fouling	0.1% caustic pH 12, 30°C followed by 0.2% hydrochloric acid, pH 2, 45°C	0.1% caustic plus 0.03% sodium dodecylsulfate, pH 11.5, 35°C
Biofouling	0.1% caustic, pH 12, 30°C	0.1% caustic plus 0.03% sodium dodecylsulfate, pH 11.5, 35°C

* CPA membranes only

a stand-alone unit, which is connected to the RO skid via removable hoses, or it can be permanently plumbed into the RO skid. A cleaning skid includes three main pieces of equipment: a tank, a recirculation pump, and a cartridge filter.

13.2.4.1 *Cleaning Tank*

The cleaning tank is typically made of plastic, but fiberglass and stainless steel versions are available. (Keep in mind chemical compatibility when selecting the type of tank to use.) The tank should be sized to handle the hold-up volume in the pressure vessels plus the volume in the piping or hoses. The approximate hold-up volume in an 8-inch diameter, 6-module long pressure vessel is about 52.2 gallons. Thus, for a 4:2-6M array, the cleaning tank would need to be about 210 gallons for the pressure vessels (since only one stage is cleaned at a time) plus piping volume for about 250 gallons total.

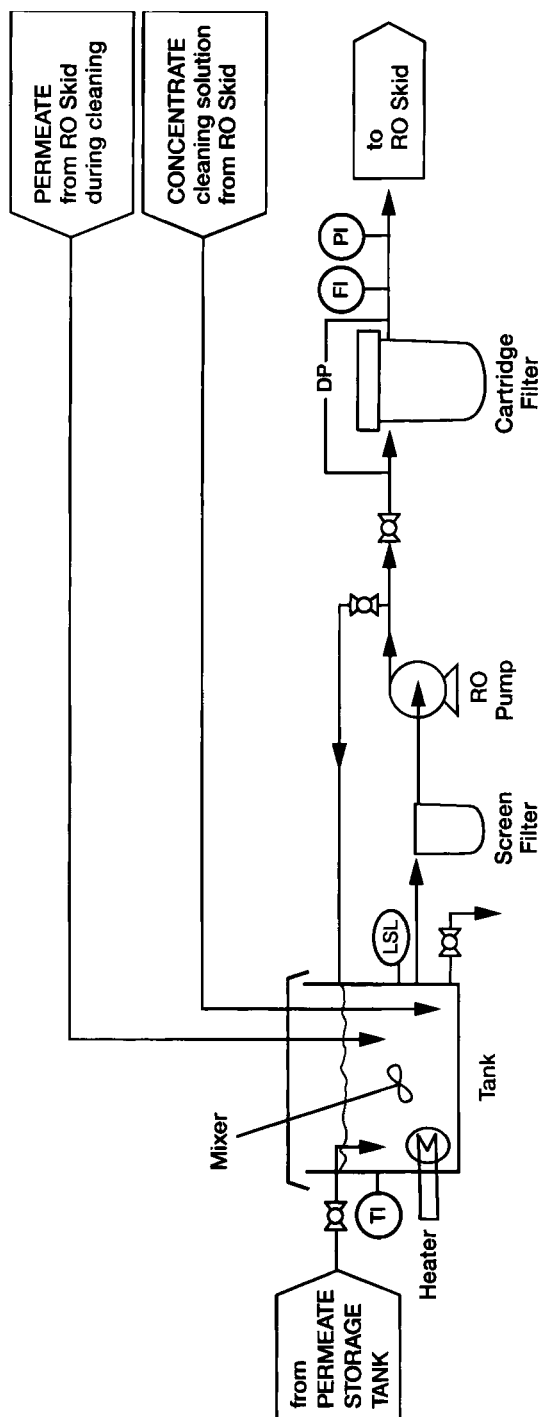


Figure 13.6 Cleaning skid schematic.

13.2.4.2 *Cleaning Recirculation Pump*

The recirculation pump is typically a centrifugal pump. It should be sized to handle the maximum recirculation rate for the largest stage in the RO skid. For example, given a 4:2-6M array, the recirculation pump must be able to deliver 45 gpm per pressure vessel or 180 gpm at less than 50 - 60 psig.

13.2.4.3 *Cartridge Filter*

The cartridge filter is typically a 5- to 10-micron nominal cutoff filter. The housing can be PVC, FRP, or SS. It should be equipped with a differential pressure gauge to measure the pressure drop across the filter. Cartridge filters should be changed out with every new cleaning solution.

13.3 **Membrane Lay-Up**

Sometimes, it is necessary to shut down an RO system for a host of reasons. For example, a peaking power facility might need to shut down an RO system for several days or months depending on the demand for power.

13.3.1 **Short-Term Lay-Up**

Membranes that are laid-up for less than two weeks should follow short-term lay-up procedures. These procedures are designed to minimize fouling, scaling, and microbial growth on the membranes. General procedures are given below.

1. Upon shut-down of the RO, the membranes should be flushed with feed water under minimal pressure (see Chapter 13.1.1) or with RO-permeate or better quality water. Flushing with higher-quality water will help remove foulants and scale and better preserve the membranes.⁹ While flushing, the air from the system should be vented.
2. When the pressure vessels are filled, the valves should be closed to prevent air from entering into the system.
3. Repeat steps 1 and 2 every 5 days.⁹

13.3.2 Long-Term Lay-up

Membranes that are left off line for extended periods of time should follow these long-term lay-up procedures. Again, these procedures are designed to minimize fouling, particularly with microbes, and scaling.

1. The RO membranes should be cleaned using the procedures outlined in Chapter 13.2.2. Flush the membranes with permeate or better quality water.
2. The cleaned membranes should then be flushed with an approved biocide, such as sodium metabisulfite (1 to 1.5 wt%) while venting air outside of the pressure vessels. Non-cobalt-catalyzed sodium metabisulfite should be used.
3. When the RO is completely filled with biocide, all valves should be closed to prevent air from deactivating the biocide. To check to see if the pressure vessels are completely filled, the solution should be allowed to overflow through an opening in the top of the RO skid that is higher than the highest pressure vessel being filled.
4. Repeat steps 2 and 3 as necessary. If sodium metabisulfite is being used, the steps should be repeated when the pH reaches 3 or lower. For other biocides, the steps should be repeated every 30 days when the temperature is below 80°F and every 15 days if the temperature is greater than 80°F.
5. When the RO system is being brought back on line, it should be flushed with low-pressure feed water for about an hour. Any permeate generated should be sent to drain. The permeate should be checked for residual biocide prior to returning the RO unit to service. If any is present, flushing should continue.

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TROUBLESHOOTING

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Troubleshooting

Despite all the pretreatment and attention to system hydraulics, most RO systems will eventually show degradation in performance. This degradation manifests itself as a loss in permeate flow, a loss of salt rejection, or an increase in pressure drop. As discussed in Chapter 12, membrane fouling, scaling, or degradation can cause these conditions. Figure 14.1 shows the relative causes of membrane fouling and scaling for 150 membranes that were autopsied.¹ As the figure shows, organic fouling is by far the leading cause of membrane fouling for the membranes autopsied. Scaling accounted for less than 5% of the total materials found on these membranes. This is because pretreatment, such as sodium softening and antiscalants, are generally effective at minimizing the formation of scale on RO membranes. Data not shown in the figure is microbial fouling. This is because all of the membranes autopsied showed some degree of microbial fouling. Conclusions in the report indicate that microbial fouling is a major problem is the counts on the membrane itself reach 10^4 CFU (colony forming units)/cm².¹

This chapter offers techniques useful in troubleshooting RO systems. The objective of troubleshooting is to identify membrane system irregularities and to investigate modes of membrane system failures, with the intent of eventually restoring membrane performance.² The ability to correct diagnose and rectify problems with an RO system is critical to keeping the unit on line.

There are two types of irregularities that an RO system can experience, those that are acute, and those that are chronic. Acute irregularities occur when there is a temporary change in feed water quality or an upset in the pretreatment system. Acute situations must be dealt with quickly before they have a chance to foul, scale, or degrade the RO membranes. Chronic problems, on the other hand, can take a long time to manifest themselves, and can result in membrane fouling, scaling, or degradation before the user can determine a definitive cause(s). Note that in most cases, there is more than one cause for performance decline.

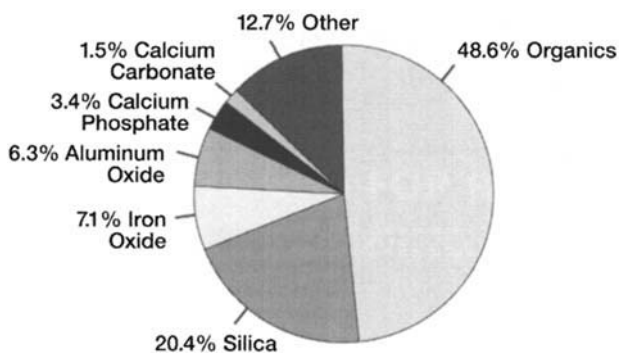


Figure 14.1 Causes of membrane fouling and scaling for 150 autopsied membranes. *Courtesy of M. Fazel and The Engineering Society of Western Pennsylvania.*

There are several investigative techniques that can be used to troubleshoot an RO system. These include:

- Mechanical evaluation
- General Performance Issues
- System design and performance projections
- Data normalization
- Water Sample testing
- Membrane integrity testing
- Profiling and probing
- Autopsy

In general, more than one of these techniques is necessary to get a complete picture of how the system is operating and to determine what might have led to the loss in performance.

14.1 Mechanical Evaluation

Perhaps the first line of investigation should be a mechanical evaluation of the system to rule out causes such as instrumentation or valves. Typical investigations include an instrument check; they should be tested and calibrated. Filters should be checked for channeling. Valves should be checked to determine if they are functioning properly. Filters and softeners should be checked to determine whether the specific flow rate or flux is within design guidelines.

Once mechanical issues have either been ruled out or identified as causes for poor performance, the performance of the RO system

should be evaluated to establish whether other factors are contributing to poor performance.

14.2 General Performance Issues

There are several general performance issues that should be investigated when troubleshooting an RO system. These include the following issues:

- *Membrane cleaning frequency:* RO systems that require more than four cleanings per year have a membrane fouling and/or scaling problem.
- *Cartridge filter replacement frequency:* high replacement rates (every 2 weeks or less) could indicate a fouling problem. Low replacement rates (every 1 month or more) could lead to microbial fouling as microbes grow in the “old” cartridges.
- *Chemical application:* dosages and controls should be checked to ensure proper feeding of chemical. This includes acid/caustic, chlorine or other oxidizer, coagulant in pretreatment, and antiscalant.
- *Visual inspection:* simple visual inspections of pretreatment and RO systems can indicate potential for fouling or scaling.
 - Check tanks and piping for mold or biogrowth.
 - Open the feed side of a pressure vessel and inspect for fouling. Biofouling will feel slippery and may also have an odor.
 - Open the concentrate side of a pressure vessel and inspect for scaling. Scale will feel rough to the touch.

14.3 System Design and Performance Projections

14.3.1 System Design

Design of an RO system has a great effect on the potential for fouling or scaling the membranes. As discussed in Chapter 9, feed water flow, concentrate flow, water flux, and recovery all affect the ability of the membranes to foul and scale. Flow rates affect the concentration polarization boundary layer where fouling and scaling occur (see

Chapter 3.4). Flux and recovery affect the concentration of contaminants within the boundary layer. Following proper design guidelines, fouling and scaling can be minimized. And, when design guidelines are disregarded, fouling and scaling are very likely to occur.

Should an RO system show signs of performance problems, the design of the system should be explored. Do the flow rates and fluxes agree with the guidelines presented in Tables 9.1, 9.2, and 9.3? If not, then the effect the variation(s) has on performance should be evaluated.

14.3.2 Performance Projections

The same performance projection software programs described in Chapter 10 for the design of RO systems can also be used to evaluate the performance of an RO system. Note that in actual application, the flow performance may vary by up to $\pm 15\%$ from nominal values (which are shown in the projections).³ Salt rejection may also vary for individual membrane modules, but will never be less than the minimum specified salt rejection in the manufacturer's literature. Thus, actual, stabilized performance may not agree with performance projections, but should be close for systems larger than about 125 gpm (assuming 8-inch diameter modules).³ For example, the actual permeate TDS should be not greater than about 1.5 times the projected concentration. Permeate flow should vary by no more than about $\pm 5\%$. If actual performance varies greatly from that projected, membrane fouling, scaling, or degradation may be occurring.

Note that there is a period after installation of new membranes, where the performance of the membranes is not stable. During this period, which can last up to 2 weeks of continuous operation, the permeability and salt passage of the membrane both decline.⁴ The decline in performance is due somewhat to compaction and is worse for seawater and wastewater applications. Other reasons for the decline are not clear but may include the degree of hydration of the membrane upon start up. Up to 10% of initial permeability and salt passage can decline during this time of destabilized performance.⁴ The decline in permeate flow is shown in Figure 14.2.

Permeate back pressure must be taken into account when evaluating system performance. If not considered while running projections, actual performance will show a higher feed pressure than what was projected. Hence, if observed operating pressure is greater than predicted, review the projection to determine if permeate

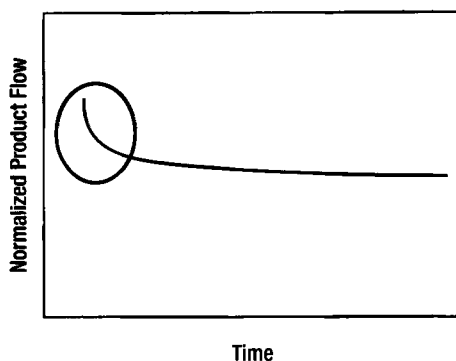


Figure 14.2 Typical permeate flow decrease upon start up of new membranes due to compaction and possibly degree of membrane hydration.

pressure was included. If permeate pressure was considered in developing the projection, then fouling or scaling of the membranes could explain the difference between actual and predicted operating pressure.

Pressure drop must also be assessed. Most projection programs assume a piping loss of about 5 psi (35 kPa) per stage plus the pressure drop through the membrane modules in the projection program. Should the actual pressure drop exceed the predicted pressure drop, there could be two explanations:

- membrane fouling or scaling, or
- location of the pressure sensors leading to misleading readings. Restrictions in the feed and concentrate headers could lead to higher than predicted pressure drop. The pressure sensors should be located as close as possible to the pressure vessel to avoid these restrictions, and away from high-turbulence areas such as valves.

14.4 Data Assessment

Data, particularly normalized data, is evaluated to determine the nature of the loss in membrane performance (see Chapter 11.3 for a complete discussion on data normalization). Normalized permeate flow, salt rejection, and differential pressure should be evaluated to determine trends in performance.

Table 14.1 is a troubleshooting matrix that shows trends in normalized performance and possible causes. The table lists different sets of performance parameters and possible causes for each. Some symptoms are the same, such as items one through three in the table, but each occurs in a different place in the RO system, and thus, each has a different cause.

As discussed in Chapter 13.2, normalized data should be used to determine when the membranes should be cleaned. Normalized data should be scanned to determine if the membranes were cleaned on time, or if potential fouling and/or scaling could become permanent because the membranes were not cleaned on time.

Table 14.1 Reverse osmosis troubleshooting matrix.

Case	Normalized Permeate Flow	Normalized Salt Passage	Normalized Differential Pressure	Location	Probable Causes
1	Up	Greatly up	Stable	1 st stage	Oxidation
2	Up	Greatly up	Stable	All stages	Hydrolysis
3	Up	Greatly up	Stable	Random	O-ring, membrane failure
4	Greatly down	Up	Up	Last stage	Scaling
5	Down	Stable*	Greatly up	All stages	Biofouling
6	Greatly down	Stable*	Stable	All stages	Organic fouling
7	Greatly down	Up	Up	1 st stage	Colloidal fouling
8	Greatly down	Greatly down	Stable	All stages	Compaction

* can increase as a secondary effect resulting from the fouling layer disrupting the cross-flow action on the membrane, which increases concentration polarization

The effectiveness of cleaning can also be determined by following the trend in normalized data. Not only should the performance after cleaning return to initial conditions, the performance should remain at such a level for an extended period of time. Figure 14.3 compares the performance of a well-cleaned RO system to one where the cleaning was less than successful. Normalized permeate flow rate returns to near initial performance and remains high for the well-cleaned RO system. The performance of the other RO system does not return to initial flow and drops off again rapidly. This could be a sign of irreversible fouling or scaling.

Normalized performance can also be used to track the improvements in pretreatment, for systems that can be successfully cleaned but performance still drops off too rapidly. Recall the case study discussed in Chapter 11.3.1.1 and the normalized permeate flow in Figure 11.3 (reproduced here). The membranes were fouled with colloids that carried over from the cold-lime softener and that passed through the three multimedia filters, which were all channeling. Iron oxide was also fouling the membranes; the iron was introduced by corrosion of the stainless steel pipe when exposed to the hydrochloric acid used to adjust the pH after the cold-lime softener. As improvements were being made to the pretreatment system, the normalized permeate flow returned to a higher value after each cleaning. Eventually, the normalized permeate flow reached the initial normalized permeate flow and stayed fairly constant without benefit of additional cleanings. Additionally, the normalized permeate flow now tracked with the actual permeate flow. Both these conditions are indicative of the improvements in pretreatment being successful.

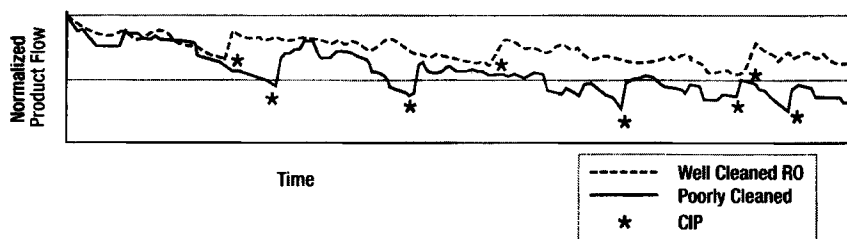


Figure 14.3 Trends in normalized data showing effectiveness of membrane cleaning. Note the decline in permeate flow during the first few days of operation due to compaction (see Chapter 14.1.2.2).

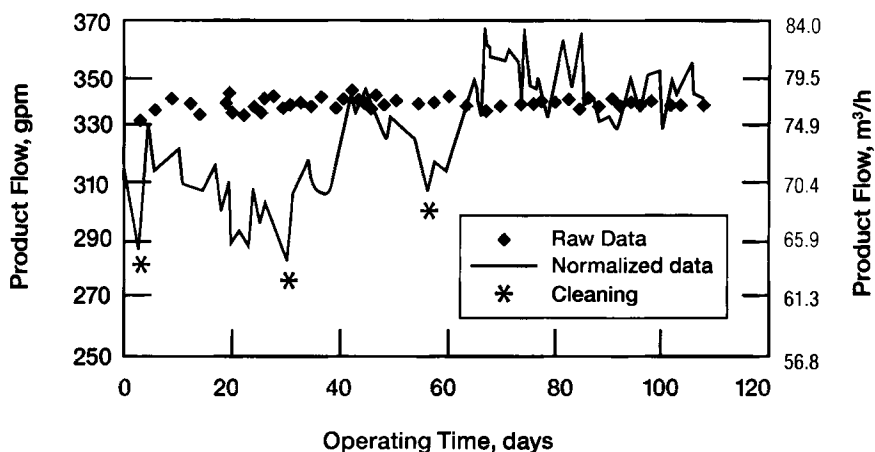


Figure 11.3 Normalize product flow rate data from a facility operating on cold lime softened Delaware River water.

14.5 Water Sampling

Water sampling cannot be under-rated as to the importance it has to the performance of the RO system. As described in Chapter 7, water source plays a big role in determining the degree of membrane fouling, scaling, and degradation that RO system might experience. When troubleshooting, the water source should be determined. If the water is municipal, the ultimate source needs to be determined. The pretreatment system should be reviewed to ensure that it meets the challenges of the feed water.

A detailed water analysis is also required. The species should be compared to Table 7.1, to determine the potential for fouling, scaling, and degradation. Specific species can be identified that may have already fouled, scaled, or degraded the membranes, as shown in Table 14.1.

A troubleshooting technique that will help in determining what species have fouled or scaled a membrane is a mass balance around the membrane. A species is selected, such as iron or aluminum, and a mass balance is taken around the RO system. The amount of that species on the feed side should be equal to that amount in the permeate plus the concentrate. Keep in mind that this is conservation of mass not concentration (there is no conservation of concentration). So, to calculation the amounts, the user will have to take the concentration and multiply by the flow rate. This will generate an amount per unit time that

will be suitable, if all time frames are the same (for example, if gallons per minute is used). If a species come up short in the concentrate plus permeate versus feed mass balance, it is probable that this species has deposited on the membranes or elsewhere in the membrane module.

14.6 Membrane Integrity Testing

If the normalized salt rejection is low or the normalized permeate flow is high, the integrity of the membrane may be in question. The vacuum decay test is a direct test for the integrity of a spiral wound RO membrane module. The test is best used to identify leaks within the membrane modules rather than leaks due to chemical attack. The test requires the isolation of an individual membrane module or the entire pressure vessel. A vacuum is then pulled on the membrane(s) and the rate of decay in pressure is observed. A decay of greater than 100 millibar per minute is indicative of a leaky membrane. Refer to ASTM Standards D3923⁵ and D6908⁶ for a more detailed review of the technique.

14.7 Profiling and Probing

Profiling and probing are two techniques typically used together to determine if and where in an RO system membranes are scaled or leaking. These techniques should be used when the conductivity in the product of an RO system is high.

Profiling is used to determine which RO pressure vessel(s) has high conductivity. To profile, the user must have access to the permeate channel on each pressure vessel in the skid(s) of interest. A hand-held meter is then used to determine the conductivity out of each pressure vessel. The conductivity is then recorded in any manner that allows for comparison of all the values form a single skid, such as a bar graph. Figure 14.4 shows such as bar graph for a hypothetical 4:2-6M array. The conductivities from first stage are compared to each other, and the higher values are singled out for vessel probing. The same procedure should be followed for the second stage. As shown in Figure 14.4, the first stage vessel exhibiting a conductivity of 3.6 should be probed, as should the second stage vessel with a conductivity of 6.6.

Probing involves determining the permeate concentration at various points within the pressure vessel. The most common points

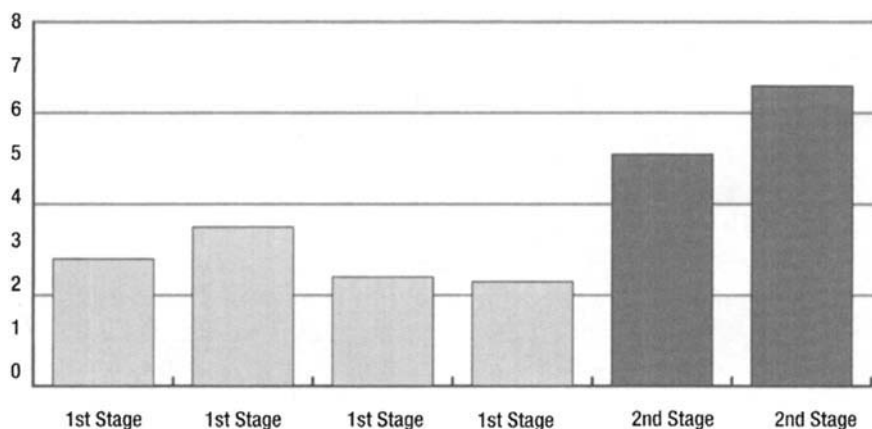


Figure 14.4 Bar graph showing permeate conductivities after profiling RO system.

selected for analysis are at the ends of the pressure vessel and at the tie point between individual membrane modules. To probe a vessel, the user must have access to the permeate port and be able to thread a long piece of Tygon®-type tubing down the permeate channels of all the membrane modules in the vessel (Tygon is a registered trademark of Saint-Gobain Corporation, Valley Forge, PA). Some RO skids are equipped with a ball valve that makes threading the tubing very easy. However, some skids are not equipped with any sort of valve and may be even hard piped, making probing difficult if not impossible. The tubing is threaded down the permeate channel of each membrane module all the way to the opposite end of the pressure vessel, as illustrated in Figure 14.5. The conductivity is measured at this point in the vessel, using a hand-held meter. The tubing is then extracted out of the pressure vessel by about 40 inches, corresponding to the length of a module, and the conductivity is measured again. This procedure is repeated until all the points of connection between modules within the pressure vessel have been measured.

Data collected during probing is plotted as a function of distance through the pressure vessel. Figure 14.6 shows what some specific problems would look like when graphed.

1. A pressure vessel with no leaks or scaling would show a slight increase in conductivity from the feed end to the concentrate end. This is a result of the feed water becoming more concentrated (and hence, the product

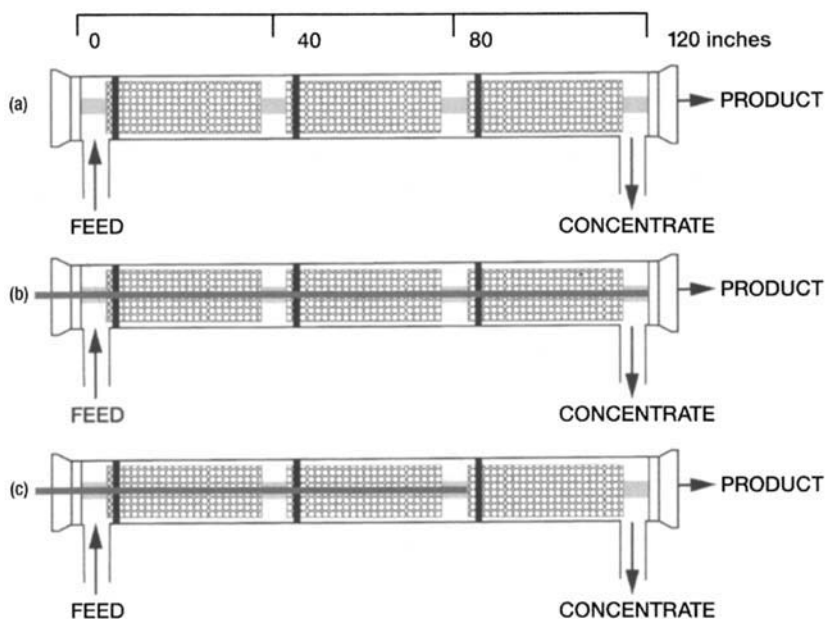


Figure 14.5 How to probe an RO pressure vessel: (a) pressure vessel prior to probing, (b) Pressure vessel with probing Tyron® tubing inserted all the way to the end, (c) pressure vessel with probing tubing partially extracted.

becoming more concentrated) as water is removed from the feed side of the membrane.

2. A pressure vessel with a leaking or missing O-ring at the feed end (or any place in the vessel) would show a spike in conductivity at that point, but then show a drop in conductivity as the distance increases away from the leak. The curve then takes on the general shape of a pressure vessel with no leaks or scale.
3. Scaled membranes would show a more gradual increase in conductivity than a pressure vessel with no leaks or scale.

Should the results of the probing be inconclusive (i.e., an increase in conductivity is noted but does not correlate with the ends of the modules), then a damaged membrane may be at fault. To test this theory, probing should be conducted again, this time measuring the conductivity every 8- to 10-inches or so. This will capture performance of individual membranes, not just the interconnectors between them.

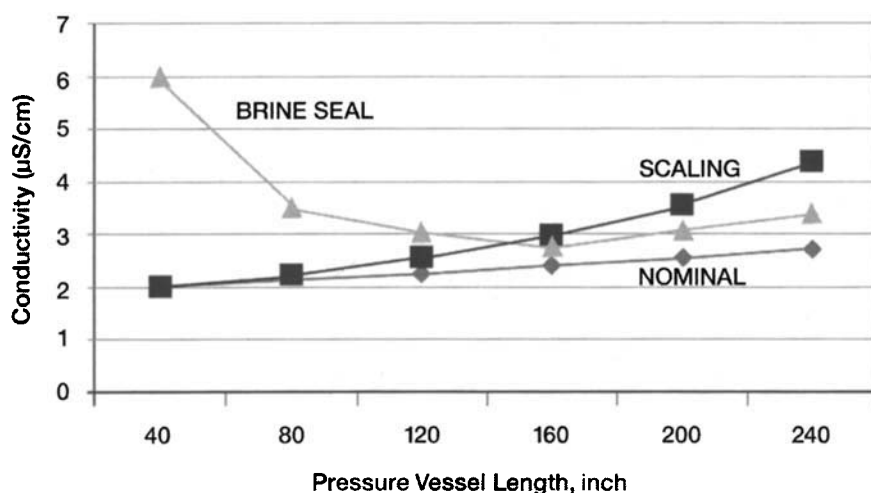


Figure 14.6 Graphed results of RO membrane probing.

14.8 Membrane Autopsy

Membrane autopsy and subsequent tests are generally the last resort in determining a definitive cause for membrane failure. This is because these tests are all destructive to the membrane module and membrane itself. The tests are used to determine the morphology of material on the membrane, the types of chemicals present, the amount of a specific species found on the membrane, and the nature of the materials that oxidized the membrane.⁷

If the user is unsure about what may be causing performance problems, it is best to autopsy at least two membranes, the lead module in the first stage of the RO system and the last module at the concentrate end of the last stage. The lead module will capture suspended solids that may be fouling the membrane while the last module will contain any scaling that may be occurring. If fouling or scaling is suspected, however, then a membrane(s) from the appropriate location in the RO system should be selected for autopsy.

Modules should be prepared for autopsy following the vendor's instructions. Typically, preparation involves bagging each module individually in plastic and boxing it into the original shipping container, if available. Modules should be sent via express mail within 2 days of removing them from service. Membranes *should not* be cleaned prior to autopsy.

14.8.1 Visual Inspection

The first inspection of a membrane submitted for autopsy is a visual one. The module itself is inspected for damage, such as crumpling or crushing damage that could have occurred due to excessive pressure drop or water hammer during operation. Figure 14.7 shows a crumpled membrane module (a) and a cracked membrane module (b). The end caps of the module are checked for damage that also may occur due to excessive pressure drop. A cracked end cap from the concentrate outlet end of an RO membrane module is shown in Figure 14.8. The feed-end cap and face of the module are studied to determine if there are any foulants present that may be blocking the

(a)



(b)



Figure 14.7 (a) Crumpled membrane module due to excessive pressure drop or water hammer (b) cracked membrane module.

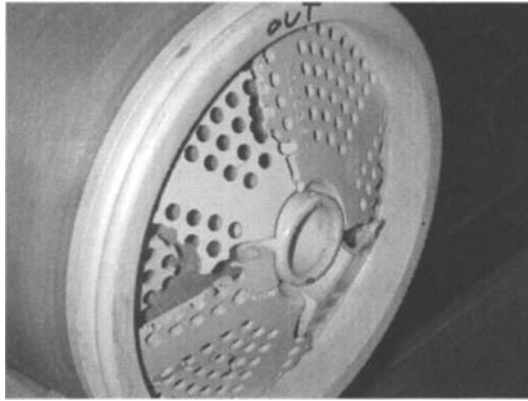


Figure 14.8 Cracked end cap as the concentrate outlet.



Figure 14.9 Feed end of membrane module showing foulants.

feed channels. Figure 14.9 shows the feed inlet end of an RO membrane module with debris blocking the feed channels to the membranes. The concentrate end of the module is viewed to determine whether any telescoping of the module has occurred (see Chapter 4.3.3). Figure 14.10a shows a severely-telescoped membrane module. In most cases, however, telescoping is not as extreme as shown in Figure 14.10a; it typically takes the form of protruding spacer and sometimes membrane sheets, as shown in Figure 14.10b.

Once the module has been opened, the surface of the membrane, glue lines, and feed spacers should be inspected for damage and

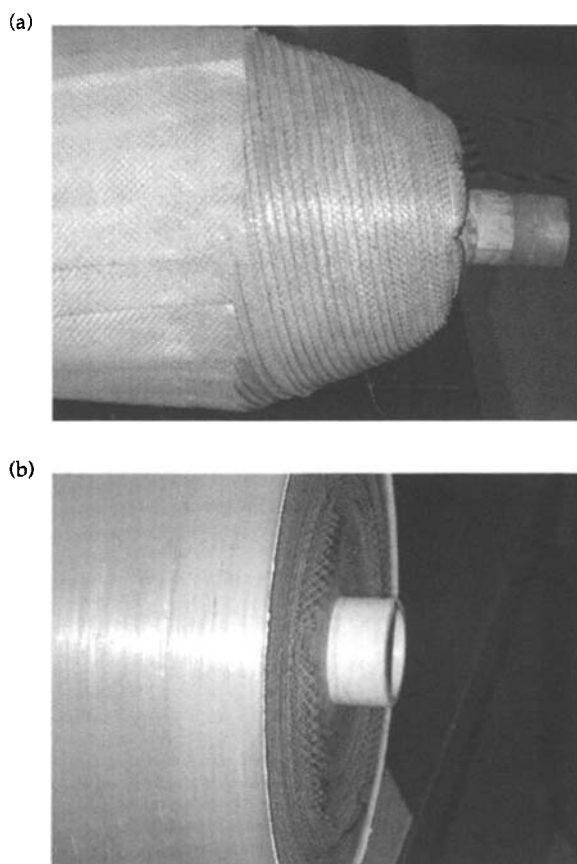


Figure 14.10 (a) excessively-telescoped membrane module due to high pressure drop resulting from considerable membrane fouling or scaling, (b) concentrate end of RO membrane module showing protruding (telescoped) feed spacer material.

for scale and foulants. Damaged glue lines can mean excessive backpressure was applied to the membrane module. Figure 14.11 shows blisters where the glue line has been breached. Permeate back pressure must never exceed the concentrate pressure by more than 15 psi, particularly at the tail end of the pressure vessel. Should such an event occur, the membranes can delaminate at or near the glue lines in the membrane module, thereby allowing feed water to mix with RO permeate water. One of the most common ways delamination occurs is when there is high pressure drop on the feed/concentrate

side of the membrane due to excessive fouling or scaling. Others ways in which delamination can occur include

- *Permeate Divert Valves:* Permeate divert valves (or any other valve on the permeate line) should be set to close only after the other valve has opened completely or the RO system has shut down. If all permeate valves were to be closed while the RO was operating, permeate would have nowhere to go and pressure would build up quickly on the permeate side of the membrane.
- *Tanks:* Tanks provide an important break in back pressure for RO permeate, if the tank is positioned properly. Permeate rarely has enough pressure to feed directly into another unit operation, so tanks provide the break in back pressure that would occur if the permeate were sent to another unit operation. However, tanks need to be carefully positioned so as not to cause significant back pressure themselves. For example, tall, narrow tanks require that permeate have enough head pressure to reach the inlet at the top of the tank. Depending on the height of the tank, this may result in permeate back pressure that is too high. Also, placement of the tank on a another level in the facility that is higher than the RO skid level can also result in high permeate back pressure. For example, placing the tank on the roof when the RO skid is on the ground floor may result in high permeate back pressure.

Other issues that may be identified upon visual inspection include:

- *General Oxidation:* Figure 14.12 shows damage to the membrane. In this case, the damage took the form of a hole in the membrane. At the same time, iron-fouled resin beads of the same size as the holes were found on the surface of that membrane (see Figure 14.13). It was deduced that the damage to the membrane in Figure 14.12 was probably caused by iron oxidation of the membrane.
- *Fouling and Scaling:* Figure 14.14a shows a feed spacer with foulants adhering to the spacer, while Figures 14.14 b and c show feed spacers virtually completely blocked with foulants and scale, respectively. This

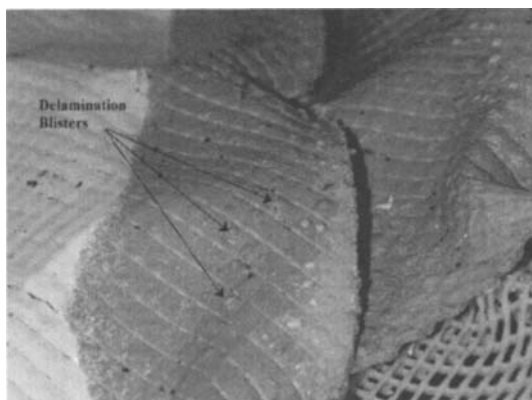


Figure 14.11 Damaged glue lines due to excessive backpressure.

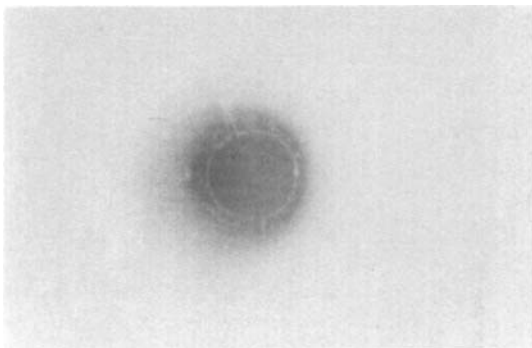


Figure 14.12 Hole in membrane surface.

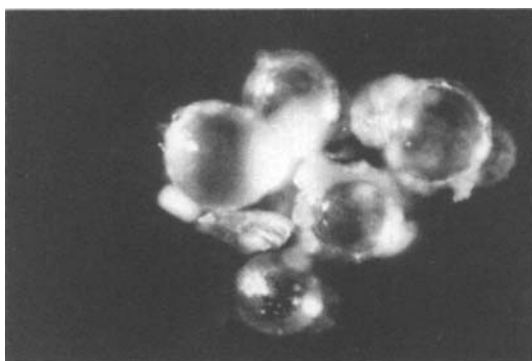


Figure 14.13 Iron-fouled resin beads on membrane surface.

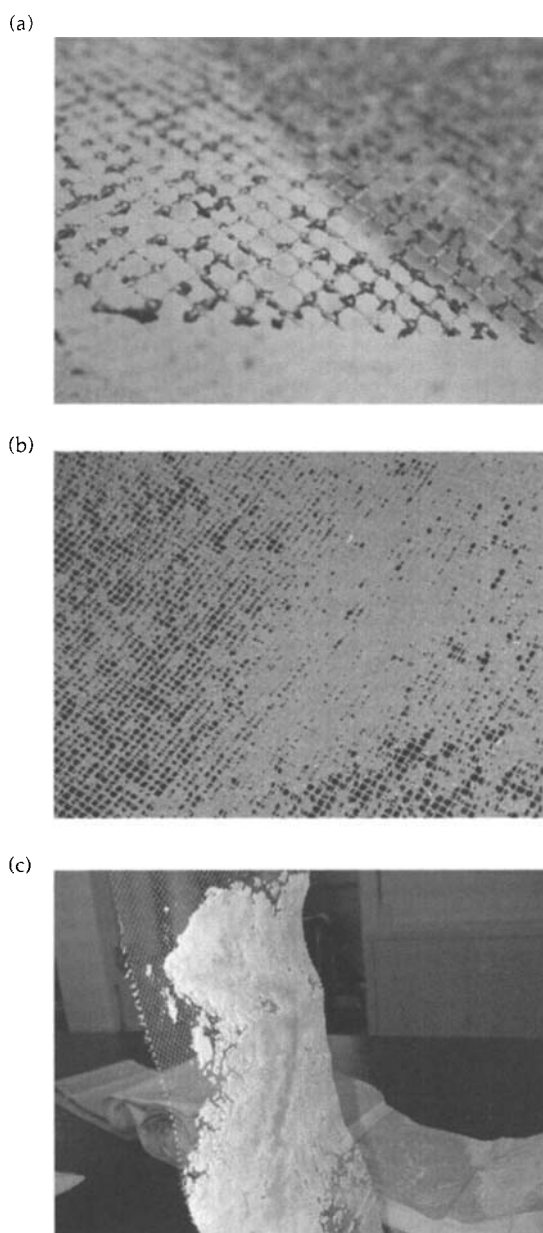


Figure 14.14 (a) foulants adhering to feed spacer; feed spacers virtually blocked with (b) foulants and (c) scale.

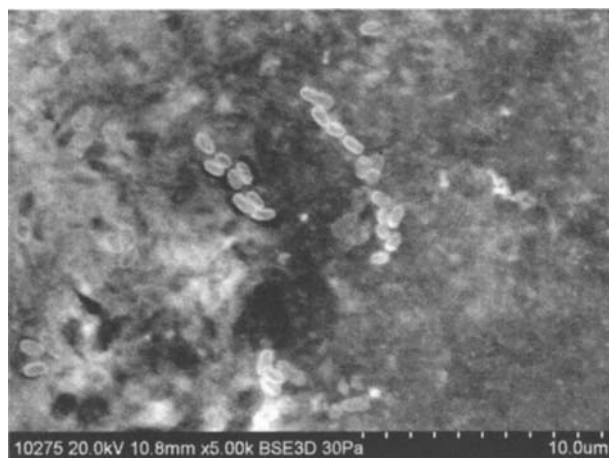


Figure 14.15 Microbial residue on the surface of a membrane (5000X view).

would seriously hamper the flow through the membrane leading to poor production, high operating pressure, and high pressure drop.

- *Microbial Fouling:* Finally, Figure 14.15 shows microbial residue on the surface of a membrane magnified 5000 times.

14.8.2 Pressure Dye Test—Rhodamine B

This test is used to determine whether a membrane has been damaged by exposure to an oxidizer or by some sort of physical attack. In the Rhodamine B test, the membrane module is pressurized with a dye prior to autopsy. If the permeate turns pink, the membrane is probably damaged. Upon autopsy, damaged areas of the membrane will be stained pink.

14.8.3 Methylene Blue Test

The Methylene Blue test is used to determine if there is any chemical or physical damage to the membrane surface. In this test, a sheet of membrane taken after autopsy is tested in a flat-sheet test apparatus. The feed side of the membrane is exposed to a 0.05% solution of methylene blue. If the membrane is damaged, the permeate will turn blue and the membrane will stain blue in the damaged areas.

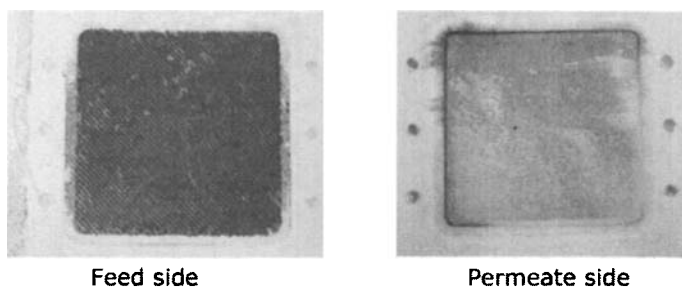


Figure 14.16 Methylene Blue test showing stained permeate side of a membrane, indicating damage to the membrane.

Figure 14.16 shows such a membrane. The darker shaded areas on the permeate side are actually stained blue from the methylene blue solution, indicating that the membrane has been damaged.

14.8.4 Fujiwara Test

The Fujiwara test is used to determine whether a membrane has been oxidized by exposure to halogens. The test measures the presence of halogenated organics in a membrane sample. The test involves a small piece of membrane placed in the bottom of a test tube. One drop of 5N sodium hydroxide solution and 2 drops of a pyridine solution are added to the test tube. The tube is then placed in a water bath at about 90°C, and held there for 30 seconds. A positive test occurs when the pyridine layer in the test tube shows a red or pink color. Note that prolonged heating of the sample will cause the color to fade or turn to brown/yellow.⁷

14.8.5 Spectroscopy

There are several spectroscopy tests that can be used to determine the nature of materials that have fouled or scaled a membrane. These are described below.

- Scanning electron microscopy (SEM), along with stereo or standard light microscopes, can be used to determine the morphology of materials on the surface of a membrane.
- Energy dispersive X-ray fluorescence (EDXRF) spectroscopy is used to determine which chemicals are

in the deposit on the membrane. This technique uses short wavelength X-rays to ionize component atoms in a material. This process emits energy in the form of a photon. The energy is characteristic of the atoms present. The term "fluorescence" is applied to this phenomenon because the absorption of higher-energy radiation results in the re-emission of lower-energy radiation.

- Inductively coupled plasma emission (ICP) spectroscopy is used to determine the amount of trace metals and silica on a membrane. This technique uses inductively coupled plasma to excite atoms and ions that emit electromagnetic radiation and wavelengths that are characteristic of a particular element. The intensity of the emission is indicative of the concentration of that element in the sample.
- Electron spectroscopy for chemical analysis (ESCA) is used to determine impurities that have been organically bound to a membrane surface. This technique measures the elemental composition, chemical state, and electronic state of elements within a material.
- Infrared spectroscopy is used to detect most organic material and some inorganic materials such as iron, silicates, carbonates, and sulfates. The technique uses the absorbance of the infrared light frequencies to detect the nature of chemical bonds present.

14.8.6 Other Tests

X-Ray diffraction can determine whether or not the material on the membrane is crystalline. This technique uses X-rays to strike the material of interest. X-rays are scattered, and from the angles and intensities of the scattered beams, the nature of the crystal can be determined.

Microbial testing for aerobic and anaerobic bacteria and fungi is typically conducted. Specific species searched for include:

- Aerobic bacteria
 - Pseudomonas
 - Spores
- Anaerobic bacteria
 - Clostridia
 - Sulfate-reducing bacteria

- Fungi
 - Molds
 - Yeasts
- Iron bacteria
- Algae: filamentous and non-filamentous
- Diatoms
- Protozoa

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6

SYSTEM ENGINEERING

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Issues Concerning System Engineering

Various engineering issues concerning RO and pre- and post-treatment are discussed in this chapter. Commonly encountered issues include:

- sodium softening and whether to place the softener before or after the RO unit
- whether to use sodium softening or antiscalant
- sizing of an RO in variable flow demand conditions
- cleaning of RO membranes on-site or off-site
- disposal of RO reject

15.1 Sodium Water Softening

Sodium softening is used to remove soluble hardness from water, including calcium, magnesium, barium, and strontium. As discussed in Chapter 8.1.6, sodium softeners are commonly used to pre-treat RO feed water to reduce the potential for scaling the membrane with hardness scales. In the next two sections, the placement of the sodium softener, either before or after the RO system, as well as the use of sodium softeners versus antiscalants are discussed.

15.1.1 Sequencing of the Sodium Softeners and RO

Traditionally, sodium softeners have been used as pretreatment to RO. Sodium softeners remove hardness and metals, such as iron and manganese, that scale, foul, or degrade RO membranes. Anecdotally, softeners are also used to help reduce suspended solids and SDI from surface or other highly fouling feed waters prior to RO. The sodium softener, in essence, acts as another barrier in front of the membrane.

The disadvantage with using sodium softening as RO pretreatment is that the softener must treat not only the permeate volume but also the volume of water that will become the reject. In other words, the softener must be large enough to treat the entire feed volume to the RO. This brings up two issues:

1. The softener system must be relatively large, as the service flow rate through a softener vessel should be about 6 – 8 gpm/ft².¹ A 500-gpm RO operating at 75% recovery (see Chapter 3.1) would require two 120-inch diameter vessel to soften the feed water and maintain the desired service flow rate while one unit is in regeneration.
2. Chloride discharge may become a concern. One 120-inch diameter vessel will generate about 3,400 gallons of 10% brine waste just from the brining step alone. A 10% brine solution contains about 6,000-ppm chloride. Current Environmental Protection Agency (EPA) guidelines call for a chronic (or continuous) chlorine discharge limit of 230 ppm to a controlled watercourse. To meet the current guidelines, the brine solution would have to be diluted by a factor of 18, or the 3,400 gallons of 10% brine would need to be diluted by 61,200 gallons of chloride-free water.

These two issues have prompted several users to move or consider moving the sodium softener from in front of the RO to after the RO, to polish the RO permeate. Post treatment of RO permeate is often necessary because the RO does not reject 100% of the hardness in the feed water. And, since the feed water is not softened, the concentration of hardness in the RO effluent will be higher than if the feed water were softened prior to the RO. Depending on the application of the permeate, polishing with a softener to remove hardness may be required.

The advantages of this configuration include the following:

1. The sodium softener is only treating the RO permeate, typically about 75% of the feed flow rate.
2. A polishing sodium softener can operate at a higher service flow rate than a primary softener. Instead of being limited to 6 – 8 gpm/ft², a polishing softener

can operate at 10 – 15 gpm/ft². The same 500-gpm RO system that requires two 120-inch diameter softener vessels for pretreatment, would only require two 84-inch diameter vessels to post-treat the RO permeate.

3. One 84-inch diameter vessel will generate about 1,700 gallons of 10% brine waste. About 44,200 gallons of chloride-free water will be required to dilute the chloride to meet discharge limits, or only about 60% of that required when the sodium softener was located in front of the RO.

The disadvantage of the post-RO arrangement is that the RO membranes are now more prone to scaling, fouling, and degradation with hardness and metals such as iron and manganese. This can be addressed by using antiscalants to minimize scaling (see Chapter 8.2.3) and appropriate filtration, such as manganese filtration, to remove iron and manganese (see Chapter 8.1.5). However, costs would have to be carefully evaluated to determine which option (softener in front of or behind the RO) is more cost effective.

15.1.2 Sodium Softening and Antiscalants

The choice of sodium softening or antiscalant is specific to each application. It is very difficult to make a blanket statement that one technology is better than the other, since both are effective at minimizing the potential for scaling RO membranes. Some designers prefer using sodium softeners because of the additional barrier they provide in front of the RO. Others prefer antiscalant to eliminate the need to handle brine waste (see Chapter 15.1.1).

For practical purposes, it generally makes sense to use sodium softeners for RO feed flow rates of less than about 20 gpm rather than conventional, flow-proportional control; difficulties in feeding based on flow control are exacerbated at low flow rates because of limited turn down on proportioning chemical feed pumps at low flow rates. However, when Nalco Company's 3D TRASAR technology is used, antiscalant can be fed to RO systems operating feed flow rates as little as 3 gpm.²

The other issue to consider is cost, both capital and operating. Capital is generally lower for the antiscalant feed system than for a softener. An antiscalant feed system can be had for as little as about \$2,500 for the pump, make-down calibration equipment, and a day

Table 15.1 Operating costs considerations for sodium softeners and antiscalant feed systems.

Operating Cost Issue	Sodium Softeners	Antiscalant Feed System
Pump Maintenance	Yes*	Yes
Pump Accrual	Yes	Yes
Vessel Maintenance	Yes	No
Tank Maintenance	Yes (brine)	Yes (day)
Salt	Yes	No
Antiscalant	No	Yes
Resin Amortization	Yes	No
Membrane Fouling Protection	Moderate**	Little
Membrane Scaling Protection	High	Moderate to High*

* for pumped-brine systems only

** Some anecdotal evidence on higher-fouling waters

+ some difficulty with calcium phosphate scale

tank for a 20-gpm RO system. A comparable softener could run about \$20,000 for a duplex system that allows for continuous operation.

Table 15.1 lists operating cost issues for sodium softeners and antiscalant feed systems. The three largest expenditures listed in Table 15.1 are the salt and resin amortization for the sodium softener options and the antiscalant itself for that option. To provide examples of these costs in greater detail, consider the following cases.

Case 1 High Hardness Well Water

Well water has a total hardness of 285 ppm as calcium carbonate, with 250 ppm calcium and 35 ppm magnesium. The remaining feed water analysis is listed below (all species listed as ppm ion):

- sodium: 75.4 • potassium: 5.7
- barium: 0.08 • strontium: 0.13

- iron: 0
- manganese: 0
- sulfate: 21
- chloride: 240
- fluoride: 0.17
- bicarbonate: 84
- nitrate: 0.3
- silica: 15
- phosphate: 0.5
- pH: 8.1

The feed is make-up water to a 110 gpm RO operating at 75% recovery. Hence, the pretreatment system must treat 150 gpm of feed water.

Sodium Softener

The sodium softener selected to treat this water is a duplex, 150-gpm system with 54-inch diameter by 72-inch side sheet vessels. Each vessel contains 50ft³ of resin. Since the duplex system operates with one vessel on line and one vessel in stand-by, the total system regenerates 2.4 times per day. At a salt dosage of 15 pounds per cubic foot, the system uses about 1,800 pounds (820 kg) of salt per day or 27.4 tons per month, assuming a 100% operating factor. At a salt cost of \$80 per ton, the total month cost for salt is about \$2,200.³

Amortization of resin typically assumes a 5-year life or replacement cycle. With two vessels at 100ft³ of resin total, the monthly amortization is \$100.

Antiscalant

Table 15.2 lists the saturation indexes for the untreated feed water, feed water with 10.2 ppm antiscalant, and 4.2 ppm of antiscalant plus 3.4 ppm sulfuric acid for pH reduction from 8.1 to 7.5. As the untreated water shows, the major species of concern are the calcium carbonate, barium sulfate, and calcium phosphate. The antiscalant does a good job with all but the calcium phosphate. To address this potential scale, acid must be added. This reduces the antiscalant demand by 60%.

Average cost for an antiscalant in 2009 runs about \$4.00 per pound.⁽⁴⁾ At a dosage of 10.2 ppm, the daily cost for antiscalant is about \$72.00 or about \$2,200 per month. With acid feed, the antiscalant requirement is about \$905 per month. Sulfuric acid, at about \$600 per ton, and 3.4 ppm feed rate, would add about \$55

Table 15.2 Saturation indexes for untreated feed water and feed water treated with antiscalant.

Feed Water Condition	Untreated	With 10.2 ppm Antiscalant	With 4.2 ppm Antiscalant and Supplemental Acid Feed
Calcium Carbonate	164	75	74
Calcium Sulfate	1.7	0.6	0.6
Barium Sulfate	200	2.4	2.4
Strontium Sulfate	0.4	0.0	0.0
Calcium Fluoride	30.1	0.0	0.0
Silica	35	23	31
Calcium Phosphate	117	103	96

per month. Monthly total for the antiscalant only option is about \$2,400, while that for the acid/antiscalant option is about \$960.

Summary

For this high-hardness case, the softener and antiscalant options appear to be about the same when only antiscalant (no acid) is used. When acid is feed in conjunction with the antiscalant, the lower cost option is the chemical one when comparing cost of salt versus cost of antiscalant/acid. Note that any local variation in cost of either salt or antiscalant can change the results. What is not accounted for is the affect a softener has on the ability to remove suspended solids and SDI out of feed water, such as any iron or manganese that may be in the well water. If the water were to contain soluble iron and manganese, the softener would remove these species and prevent both fouling and degradation of the RO membrane, thereby reducing membrane O&M costs via reducing membrane cleaning and replacement. The chemical option would require an iron filter, thereby increasing the costs of this option. If the softener were to have a positive affect on membrane performance, this cost would have to be taken into account as well. See Case 3 below.

Case 2 Low Hardness Surface Water

Surface water had a total hardness of 105 pm as calcium carbonate, with 95 ppm calcium and 10 ppm magnesium. The remaining analysis is listed below (all species listed a ppm ion):

- | | |
|------------------|-------------------|
| • sodium: 19 | • potassium: 3 |
| • barium: 0.0 | • strontium: 0.0 |
| • iron: 0 | • manganese: 0 |
| • sulfate: 46 | • chloride: 33 |
| • fluoride: 0.82 | • bicarbonate: 42 |
| • nitrate: 12 | • silica: 5.5 |
| • phosphate: 0.0 | • pH: 8.5 |

As with Case 1, the RO system produces 110 gpm at 75% recovery that requires 150 gpm of feed water.

Sodium Softener

The sodium softener selected to treat this water is a duplex, 150-gpm system with 54-inch diameter by 72-inch side sheet vessels. Each vessel contains 50ft³ of resin and the system regenerates 0.9 times per day. At a salt dosage of 15 pounds per cubic foot, the system uses about 675 pounds of salt per day, assuming a 100% operating factor. At a salt cost of \$80 per ton, the total monthly cost for salt is about \$820.

Amortization of resin typically assumes a 5-year life or replacement cycle. Given two vessels with 100 ft³ of resin total, the monthly amortization is \$100.

Antiscalant

Projections shows that calcium carbonate and calcium fluoride, with saturation indexes of 158% and 200%, respectively, are the species to be concerned with. The addition of 2.6 ppm of antiscalant would bring down the saturation indexes to 82% and 0% for calcium carbonate and calcium fluoride, respectively. At \$4.00 per pound of antiscalant, the cost for antiscalant is about \$19 per day, or about \$570 per month.

Summary

For this low hardness, surface water case, the antiscalant operating cost is lower than that for the softener. Again, the affect of the softener on improved membrane performance is not known and should be considered for this surface water source.

Case 3 Well Water with Iron and Manganese

This case is taken from an actual analysis conducted for an O&M contract for a facility in Venezuela. The system is to take potable well water from the city and generate boiler make-up water. The water has 77 ppm (as calcium carbonate) total hardness, with 60 ppm calcium and 17 ppm magnesium. The water contains 0.5 ppm iron and 0.4 ppm manganese. The remainder of the analysis is listed below (all species listed as ppm ion):

- | | |
|------------------|-------------------|
| • sodium: 17 | • potassium: 4.8 |
| • barium: <0.4 | • strontium: 0.12 |
| • sulfate: 30 | • chloride: 56 |
| • fluoride: NA | • bicarbonate: 40 |
| • nitrate: <0.16 | • silica: 11 |
| • phosphate: NA | • pH: 6.8 |

Sodium Softener

The softener is a 48-inch diameter unit with 35ft³ of resin. The softener will regenerate once per day with 15lb/ft³ of salt. At a salt cost of \$80 per ton, the sodium softener would require about \$638 per month to operate.

Antiscalant

Projections indicate that about 3 ppm of antiscalant is required to minimize scaling due to hardness. At a local cost of \$4.50 per pound, the monthly cost for antiscalant is \$46.

Table 15.3 lists the pertinent operating costs for this system. As the table shows, the cleaning and membrane replacement costs for

Table 15.3 Projected operating costs for an RO pretreatment system with and without sodium softener.

Concern	With Sodium Softener			Without Sodium Softener		
	Frequency/Dose	Cost Each	Cost per Year	Frequency/Dose	Cost Each	Cost per Year
Antiscalant	0 ppm	\$0	\$0	3 ppm	\$4.50/lb	550
Salt	1 regeneration /day	\$21* /day	\$7,665	0 regenerations	\$0	\$0
Membranes	3 year life	\$600	\$7,200	2.5 year life	\$600	\$8,640
Membrane Cleaning	4 per year	\$3,240	\$12,960	6 per year	\$3,240	\$19,440
Total			\$27,825			\$28,630

* 15 lb./ft³ dosage, \$80/ton

the two options are different. The iron and manganese make the difference in this case. A softener will remove these species from the RO feed water, thereby protecting the membranes from degradation and fouling. This results in fewer membrane cleanings and longer membrane life.

The significance of these three cases are to show that each application requires full accounting of operating costs when considering whether to use a sodium softener or antiscalant. A complete water analysis is required on which to base the operating cost assumptions (see Chapter 7). The presence of iron and/or manganese can significantly affect the design, performance, and cost to operate an RO system with pretreatment. Additionally, local pricing for commodities will also affect the cost to operate the system. Focusing on just one operating cost variable, such as the cost of salt, may not give a complete picture of the total affect the softener might have on performance of the RO and pretreatment system.

15.2 Reverse Osmosis Sizing and Capacity

Proper sizing of an RO system, particularly when the demand for product is variable, can be a challenge. Variable product demand can involve actual swings in demand or it can be brought on by low-level and high-level sensors located in product tanks that cycle the RO feed pump on and off. In either case, the ideal condition is to keep all RO skids operating continuously. Idling the membranes makes them more susceptible to fouling and scaling, especially when a shut-down flush is not employed (see Chapter 13.1.1). Furthermore, repeated start-ups subjects the membranes to physical stress and perhaps even water hammer; soft-start motors can minimize this (see Chapter 6.2).

The best approach to meeting variable product demand is to design the RO system for the average flow rate required. For example, if demand for product water is 500 gpm for 10% of the time and 350 gpm for 90% of the time, the RO should be designed for about 365 gpm. A product tank that is large enough to hold the excess 15 gpm for 90% of the time is required. Whether an RO system is on line or not is often a function of the level settings in the RO permeate or product tank. At the low set point, the RO feed pump trips on, and at the high set point, it trips off. The key is to position the level set points far enough apart that the RO system remains on line for the maximum amount of time.

Should a large tank not be an option, the alternative is multiple skids, some of which will sit idle. The key is to rotate skids on and off so that no one skid experiences most of the down time start-ups. It is also recommended that an off-line or shut-down flush be employed, and that the motor be equipped with a soft start. For the example above, two 350-gpm skids would be required. Both skids would be on line during the 500-gpm draw. A product tank with enough capacity to handle the 200-gpm overage would be required.

If demand is continuously variable, the best option is still designing for the average flow rate, as described above. However, in some cases variation in demand may be such that it still may become necessary to cycle the feed pumps on and off with level in the product tank.

15.3 Membrane Cleaning: On-Site versus Off-Site

Chapter 13.2 discussed techniques and chemicals used for cleaning of RO membranes. This section discusses the merits of on-site verses off-site membrane cleaning. Table 15.4 summarizes the advantages and limitations of on-site and off-site membrane cleaning.

15.3.1 Off-Site Membrane Cleaning

Off-site membrane cleaning involves removing membrane modules from the pressure vessels and shipping them off site for cleaning

Table 15.4 Merits of on-site versus off-site membrane cleaning.

Off-Site Membrane Cleaning		On-Site Membrane Cleaning (CIP)	
Advantages	Limitations	Advantages	Limitations
<ul style="list-style-type: none">• Expert service• More effective cleanings• Documented results	<ul style="list-style-type: none">• Higher cost• Requires second set of membranes	<ul style="list-style-type: none">• Membranes cleaned in situ—no need for replacement membranes• Faster cleaning• Less expensive	<ul style="list-style-type: none">• Less efficient• Capital outlay for cleaning skid• Storage and handling of chemicals and waste

by a 3rd party. If the RO must remain in operation, a second set of membranes is used to replace those sent out for cleaning (one of the shortcomings of off-site cleaning).

Advantages of off-site cleaning are detailed below.

- Off-site cleaning offers expert service. Personnel specifically trained for that purpose clean membranes. Experience goes a long way when it comes to membrane cleaning, because cleaning can be as much an art as it is a science.
- Off-site cleaning is generally more efficient. Off-site cleaning operations have a variety of cleaners at their disposal to use for most types of foulants and scale. If one cleaning is not successful, another cleaner(s) can be employed to improve upon results.
- Membrane manufacturers at times give special variances to off-site cleaning operations to use conditions outside the normal cleaning recommendations for membranes, as listed in the membrane specifications. For example, higher temperature and pH may be used to address biofilms and lower pH and higher temperature may be used to remove calcium carbonate scale (refer to Table 13.1).
- Results are documented. Documentation typically includes performance testing prior to and after cleaning, and comparison of the results with specifications for that specific membrane make.

Shortcomings of off-site cleaning are as follows:

- Off-site cleaning costs more than on-site cleaning. Quotes should be obtained from specific vendors, but pricing can be as high as \$150 or greater per membrane 8-inch diameter module.
- A second set of membranes is required for continued operation of the RO.

15.3.2 On-Site Membrane Cleaning

The techniques and chemicals described in Chapter 13.2 apply directly to on-site or Clean-In-Place (CIP) membrane cleaning.

Advantages of CIP are as follows.

- Cleaning is conducted with the membrane modules in situ; there is no need for a second set of membranes.
- Cleaning is faster with CIP than with off-site cleaning. A two-stage, 500-gpm RO skid can be cleaned in about 2 days. Off-site cleaning can take a couple of weeks to turn around the membranes.
- CIP is less expensive than off-site cleaning. Depending on the chemical(s) used, an 8-inch diameter membrane module can be cleaned for anywhere from \$5 to \$25 each.

Shortcomings of CIP are listed below.

- Cleaning is less efficient than with off-site cleaning. Typically, cleaning operations are limited to one cleaner per pH (e.g., one high-, moderate-, and low-pH cleaner) on site. It is too costly to stock all cleaners that "might" be needed. Additionally, membrane manufacturers will generally require adherence to cleaning specifications.
- Initial capital outlay will need to include a cleaning skid.
- On-site personnel must store and handle the cleaning chemicals as well as the wastewater that is generated.

15.4 Reverse Osmosis Reject Disposal Options

While there may be many options to deal with RO reject, each one has advantages and limitations. Some offer cost savings, while others may increase costs, but eliminate the need to dispose of a relatively large waste stream. Reject is considered non-hazardous, and contains only those constituents that were present in the feed water (only more concentrated, which can be a problem) and any antiscalants, biocides (such as DBNPA—see Chapter 8.2.5.2), or any other chemical added as pretreatment. Disposal techniques over the years have included:

- Surface water discharge
- Land application
- Sewer

- Deep well injection
- Evaporation pond

The three most common techniques for dealing with RO reject waste are discussed here.

15.4.1 Discharge to Drain or Sewer

Perhaps the most common disposal method for RO reject is simply sending it down the drain to sewer or waste treatment. However, discharging in this manner may actually not be as simple as it sounds. There are regulations and permits that may be required that limit discharge of specific contaminants.

Discharge to a natural watershed generally requires a permit that must be periodically renewed. Permit requirements vary greatly from location to location and take into account the impact discharge will have on the local ecosystem. Generally, discharge to a larger body of water results in less stringent permits, while discharge to a smaller or more delicate ecosystem, typically will encounter tighter permitting. For example, Louisiana does not regulate chlorides or hardness into the Mississippi River near the Gulf of Mexico, so sodium softener brine waste and cold-lime softener sludge may be discharged directly to the river.⁵ On the other hand, New Jersey has such stringent regulations and permits, that zero-liquid-discharge (ZLD) is often necessary.⁵

Discharge to a sewer or publicly owned treatment system (POTW) is not as regulated as discharge to a natural watershed. Most jurisdictions do not require permits and regulate only pH. However, there may be an up-charge for high concentrations of certain contaminants such as chlorides.

15.4.2 Discharge to Cooling Tower

Sending RO reject to the cooling tower is the second most common method of disposing of the reject. The concerns with this method of handling of RO reject are 1) scaling in the heat exchange equipment, 2) tower chemistry, and 3) logistics.

Calcium, and to a somewhat lesser extent, silica, can cause scaling problems for the cooling tower, particularly if the RO reject is a major portion of the cooling tower make-up. Scaling of the heat exchanger equipment results in a loss in productivity on the process side of the

facility. New chemistries that are better able to handle high concentrations of scale-forming species are being introduced, but there will always be a limit as to what the cooling system can tolerate.

The effect of the RO reject on the cooling tower chemical program is also a concern. Because the reject contains the same species as the cooling tower make-up (only cycled up 3 to 5 times), and a possible antiscalant or biocide, adding reject to the cooling tower should not adversely affect the tower chemical program, if the reject is not a large percentage of the total make-up. However, when the RO reject comprises a large percentage of the total make-up to the tower, these issues become important.

Location of the RO system relative to the cooling towers is a concern. In most facilities, the RO is not located near the major cooling towers. There may be a small, local tower located nearby, but discharging to this tower may be difficult as the RO reject could be a major portion of the make-up flow to the tower. Pipelines must be run from the RO to the major cooling towers to allow discharge of the RO reject to these towers.

Prior to making the decision to discharge RO reject to the cooling tower, an analysis should be conducted to determine what impact the reject will have on tower operations. There is a need to balance conservation and recovery of RO reject water with the impact on the cooling system.

15.4.3 Zero Liquid Discharge

The use of RO to treat and recover reject from another RO system is becoming more common, particularly in ZLD applications. A secondary RO is used to treat the reject from the primary RO to reduce the thermal evaporative requirements of the ZLD system.

Figure 15.1 shows a ZLD system with RO as a primary and secondary treatment for wastewater. As the figure shows, wastewater from a variety of sources, including cooling tower and boiler blow downs, boiler feed water makeup treatment wastes, and process wastes, is sent to the pretreatment system. This pretreatment system typically consists of a cold lime softener, either in traditional form or using microfiltration. If the wastewater is high in organics, additional treatments might include digestion (aerobic and/or anaerobic), dissolved air flotation, and stripping. Waste from the pretreatment system is sent to a sludge press, which, in turn, yields cake solids and another wastewater stream that is recycled to the

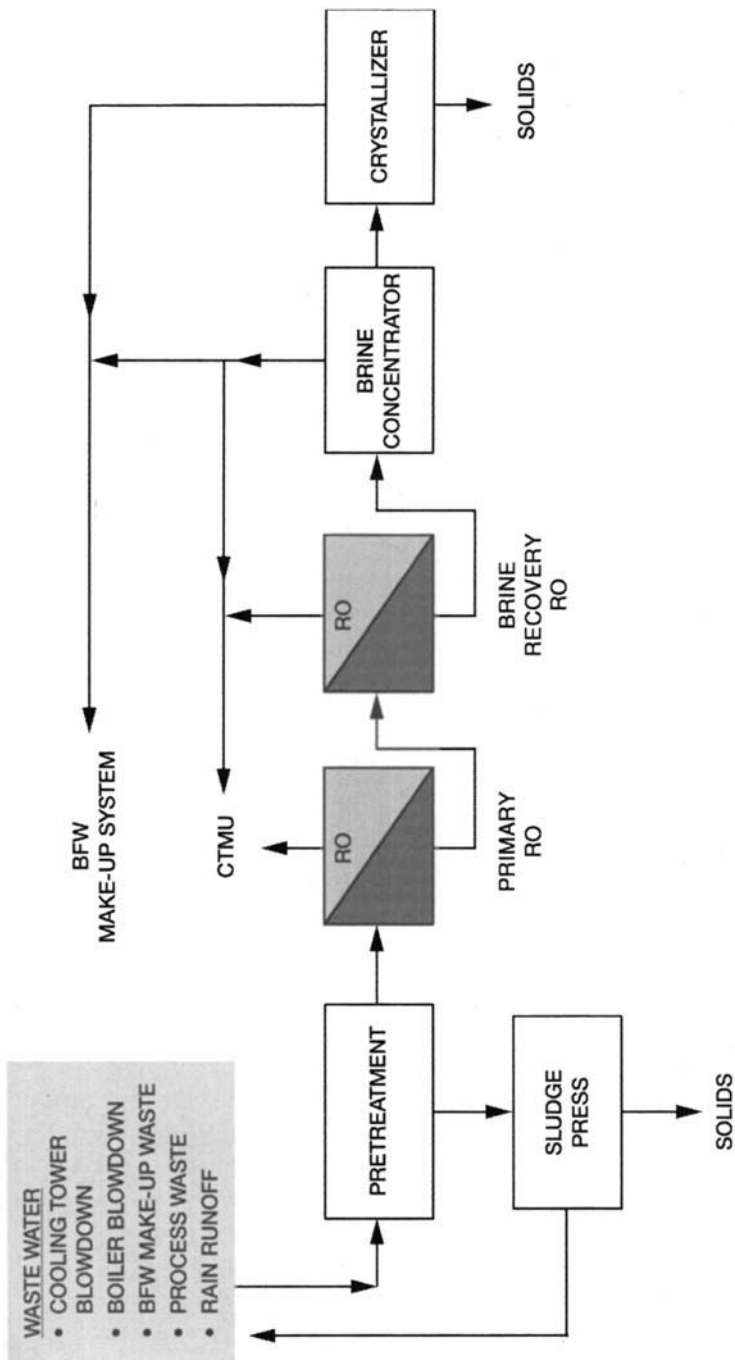


Figure 15.1 Zero-liquid discharge (ZLD) system showing primary and brine recovery RO systems.

headworks of the ZLD system. The effluent from the pretreatment system is sent to an RO, the product of which can be sent back to the cooling tower. Reject from the primary RO is sent to a concentrate or secondary RO. The secondary RO is typically a seawater system, capable of operating in the range of 45,000 to 85,000 ppm TDS.⁶ Reject from the secondary RO is sent to the thermal system, consisting of a brine concentrator and a crystallizer. The brine concentrator can achieve a concentration of about 300,000 ppm; the crystallizer yields salt solids. Using the secondary RO reduces the load on the brine concentrator by about 20%. This translates into a corresponding 20% energy savings for a single-effect brine concentrator. Capital cost of the system is also significantly reduced, as the size of the brine concentrator, one of the most expensive pieces of capital equipment in the process, can be smaller. (Refer to Chapter 16.4 to see discussion on how high efficiency reverse osmosis can be used in the ZLD system).

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Impact of Other Membrane Technologies

In this chapter, the impact of other membrane technologies on the operation of RO systems is discussed. Technologies considered include microfiltration (MF), ultrafiltration (UF), and nanofiltration (NF) as pretreatment to RO, and continuous electrodeionization (CEDI) as post-treatment to RO. This chapter also describes the HERO™ (high efficiency RO—Debasish Mukhopadhyay patent holder, 1999) process used to generate high purity water from water that is difficult to treat, such as water containing high concentrations of silica.

16.1 Microfiltration and Ultrafiltration

Microfiltration and UF are pressure-driven membrane separation technologies that use positive pressure or suction to separate particles (and high molecular-weight soluble species in the case of UF) from solution (see Figure 1.1). The history of MF and UF membranes goes back to the early 1900's. Bechhold developed the first synthetic UF membranes made from nitrocellulose in 1907.¹ He is also credited with coining the term "ultrafilter." By the 1920's and 1930's both MF and UF nitrocellulose membranes were commercially available for laboratory use. The first industrial applications of MF and UF came in the 1960's and 1970's. Microfiltration membranes became viable for industrial application in the 1970's when Gelman introduced the pleated MF cartridge.¹ Ultrafiltration membranes become industrially viable in the 1960's when Amicon began preparing UF membranes using a modified Loeb-Sourirajan method (see Chapter 4.2.1).¹

Microporous membranes are used to effect the separation by MF and UF processes. These microporous membranes differ from polyamide composite RO membranes in that they are not composites of two different polymeric materials; they are usually constructed using a single membrane polymeric material. In simple terms, both UF and MF technologies rely on size as the primary factor determining which

suspended solids and high molecular-weight dissolved organics are retained by the respective membranes. Due to the microporous nature of MF and UF membranes, the pressure required to drive water through them is significantly lower than for RO membranes, typically less than 100 psi; initial clean membrane differential pressures can be as low as 1.5 to 2.0 psi, depending on the membrane material.

Common polymers currently used to prepare MF and UF membranes include:

- Polypropylene (PP): a hydrophobic membrane with good chemical resistance and tolerance of moderately high temperatures; sensitive to chlorine.
- Polytetrafluoroethylene (PTFE) (MF only): an extremely hydrophobic membrane, with high tolerance of acids, alkalis, and solvents; can be used at temperatures up to 260°C.
- Polyvinylidene fluoride (PVDF): a hydrophobic membrane (can be surface-modified to become more hydrophilic) with good resistance to chlorine. Stable pH range up to 10.
- Polysulfone (PS): good resistance to chlorine and aliphatic hydrocarbons (not compatible with aromatic hydrocarbons, ketones, ethers, and esters), stable pH range from 1 – 13, and tolerance of up to 125°C.
- Polyethersulfone (PES): same as PS membranes. Stable pH range up to 11.
- Polyethylene (PE): poor resistant to chemical attack and relatively low strength.
- Polyacrylonitrile (PAN): good chemical resistance, but resistant to flexing.
- Cellulose triacetate (CTA): Polymer with complete acetylation of cellulose that is hydrophilic, but suffers from very narrow acceptable temperature (<30°C), and pH (4-6 nominal, 2 – 9 occasional) ranges. This material is also highly susceptible to microbial attack. (In-filco Degremont is the only manufacatuer using this material as of this publication.)

Figure 16.1 compares the strength and elongation characteristics of various polymers used in the manufacture of MF & UF membranes. High strength and good elongation characteristics are important

for the integrity of the membrane. Both properties ensure that the membrane material, particularly in the form of hollow fibers (see discussion below) can flex without breaking. Flexing is an important property for hollow fiber UF and MF membranes, which have a tendency to move in the turbulence of the flow of feed water.

Wettability or hydrophilic properties of UF and MF membranes is also important characteristics. Wettability is measured using contact angle. Contact angle is the result of the interface/surface tensions between liquid and solid. Consider a liquid droplet at rest on a flat, solid surface, as shown in Figure 16.2. The contact angle, Θ , is the angle formed by the solids surface and the tangent line to the upper surface at the end point. A smaller contact angle means the wettability and hydrophilicity of the solid material is greater. Table 16.1 lists the contact angle

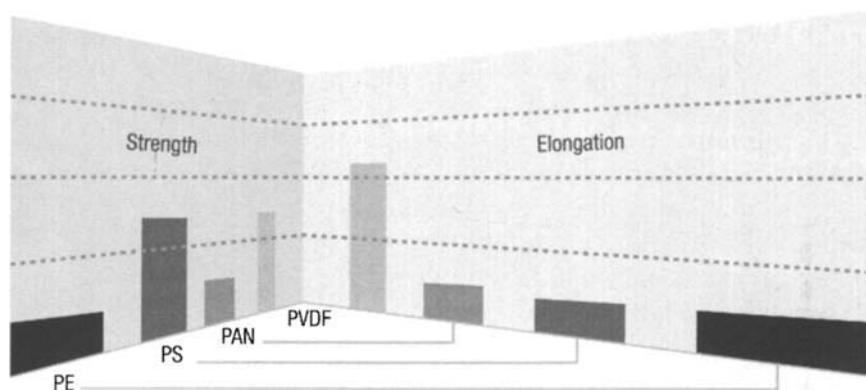


Figure 16.1 Strength and elongation characteristics of various UF and MF polymers. *Courtesy Dow Water and Process Solutions.*

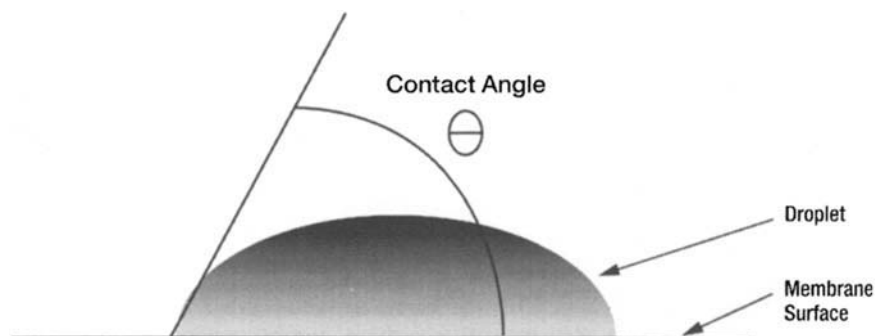


Figure 16.2 Contact angle between a liquid droplet and a flat, solid surface.

Table 16.1 Contact angle for common membrane polymers.

Polymer	Contact angle (degrees)	
PES	44	HYDROPHILIC
PAN	46	
PVDF	66	
PP	108	
PTFE	112	HYDROPHOBIC

for some common polymers. A hydrophilic membrane is usually preferable over a hydrophobic membrane for UF and MF applications.

Microfiltration and UF membranes can be asymmetric, with a denser side and a more open side, or uniform without macrovoids (See Figure 16.3). The open area behind the denser surface in an asymmetric design means there is less resistance to water permeating the membrane. Operating pressure can be lower and the membrane systems can be more productive. The limitation of the asymmetric design is that the material, predominately used in the hollow fiber configuration, is not as strong as the uniform cross section.

Microfiltration and UF membranes are available in tubular, spiral wound, and hollow fiber membrane module configurations. Tubular and spiral MF and UF modules are similar to RO tubular and spiral wound membrane modules described in Chapters 4.3.2 and 4.3.3. However, while the thickest feed spacer in a spiral RO module is 34-mil, UF and MF modules nominally have up to a 45-mil spacer due to the relatively high concentration of suspended solids these membranes are called upon to treat (TriSep™ Corporation offers a special 65-mil spacer for dairy applications).

Improvements made over the last few years in MF and UF membranes and modules, including the development of a new generation of hollow-fiber (HF) membranes and modules for industrial applications has led to wider application of these membrane separation technologies.² The new generation HF membranes are characterized by high porosity, strength, and flexibility, all important characteristics for MF and UF applications.

Microfiltration and UF hollow fiber membranes are different than the hollow fine fibers discussed in Chapter 4.3.4. The MF and UF membranes are thicker and not quite as flexible, resembling fine-diameter straws rather than human hair. Diameter of fibers ranges

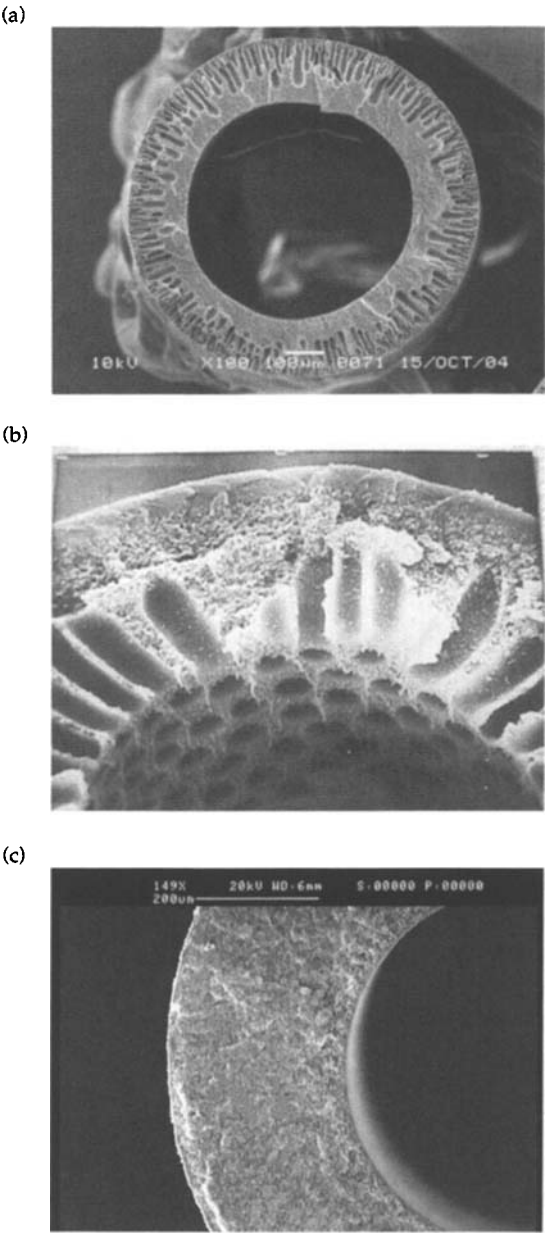


Figure 16.3 Asymmetric (inside-skinned (a) and outside-skinned (b)) and uniform (c) cross-section UF membranes. *Figure c courtesy of Dow Water and Process Solutions.*

from about 200 to 3,000 microns with a cross-sectional membrane thickness of 100 – 200 microns. The fibers are not folded into the pressure vessel as they are with hollow fine fibers, but are usually potted and open on both ends. (Note that Koch's Puron® submerged hollow fiber membrane is "potted" at one end only, with the other end unencumbered and sealed (Puron® is a registered trademark of Koch Membrane Systems, Wilmington, MA). The membrane module assembly in operation resembles a "weed" blowing in the wind. This configuration is less prone to fouling on the outside of the fibers at the sealed end than a hollow fiber that has both ended potted).

Hollow fibers can be created with the dense side on the inside or lumen of the fiber or on the outside of the fiber, or they can be double-skinned, where both the lumen and the outside of the fiber are dense. Location of the denser side of the membrane determines whether the service flow is outside-in or inside-out. Outside-in systems are typically used in a dead-end mode (or some variation thereof), while

Table 16.2 Advantages and limitations of outside-in and inside-out service flow designs for hollow fiber MF and UF membranes.

Service Flow	Outside-In	Inside-Out
Denser Side of Membrane	Outside of the fiber	Inside of the fiber
<ul style="list-style-type: none"> Advantages 	<ul style="list-style-type: none"> Minimal filtration pretreatment Easier membrane cleaning Higher surface area means more filtration area per fiber 	<ul style="list-style-type: none"> True cross flow velocity minimizes concentration polarization and membrane fouling When no air is utilized for backwashing the less fiber movement leads to breakage of fewer fibers Can be created with a variety of inside fiber diameters
<ul style="list-style-type: none"> Limitations 	<ul style="list-style-type: none"> Solids collect around and at potted ends of fibers Requires air scour to clean Cannot perform true recycle 	<ul style="list-style-type: none"> High pressure drop limits fiber length Requires significant filtration for pretreatment

inside-out systems are typically dead-end, but can also provide true cross flow. Manufacturers of outside-in membranes include Siemens (Memcor®), GE (ZeeWeed®), and Pall (Aria™--Aria is a trademark of Asahi Kasei Corporation, Tokyo, Japan). Koch manufactures inside-out membranes (Romicon®), as does Norit (X-Flow). Some hollow fiber membranes, such as Dow Water and Process Solutions' UF membrane, are skinned on both in inside and outside of the fiber, giving the fiber more strength thereby minimizing fiber breakage (see Figure 16.3c). The Dow membranes are operated in an outside-in flow pattern. Table 16.2 lists the advantages and limitations of outside-in and inside-out service flow designs.

Tubular and spiral configuration have advantages and limitations as well. Tubular membranes, with diameters of ½ to 1-inch can handle high concentrations of suspended solids; the chance of plugging the tubes is minimal. Fouled tubular membranes are easily cleaned mechanically using a sponge ball to scour the membrane surface free of foulants. However, the packing density of tubular membranes is not high, so this configuration requires a lot of floor space. Spirals, on the other hand, require less floor space, but require pre-filtration to minimize plugging of the feed channel spacer with

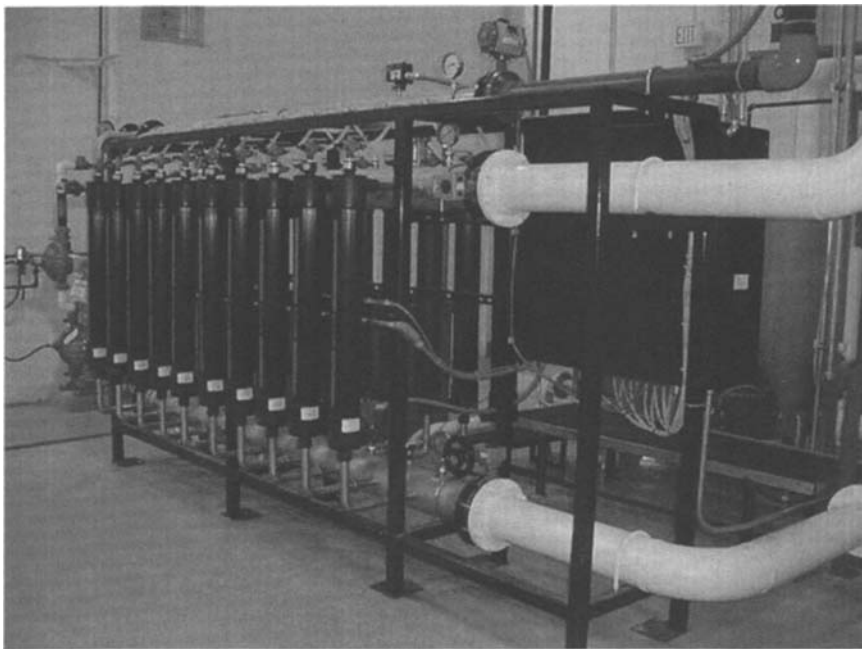


Figure 16.4 Spiral wound ultrafiltration system. *Courtesy of Koch Membrane Systems.*



Figure 16.5 Hollow fiber microfiltration system. *Courtesy of Koch Membrane Systems.*

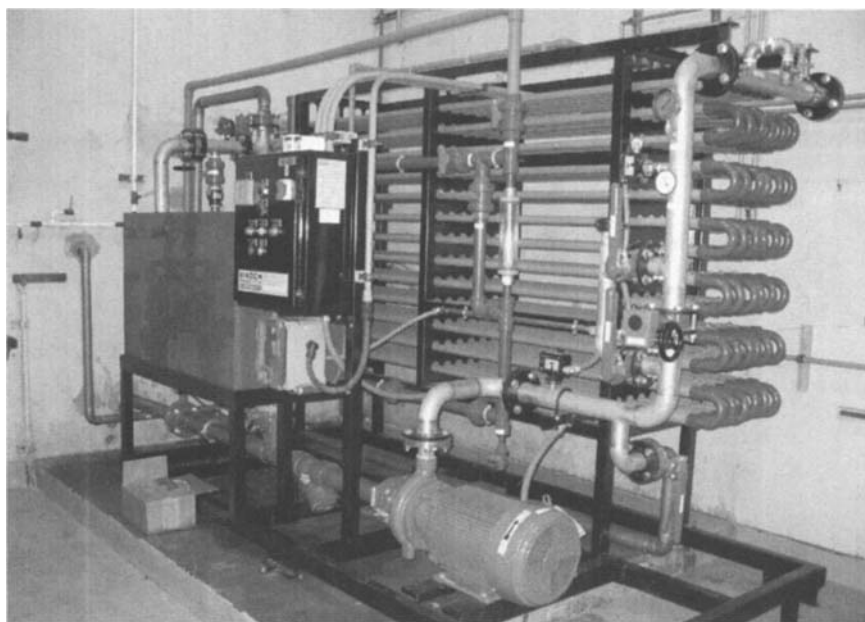


Figure 16.6 Tubular ultrafiltration system. *Courtesy of Koch Membrane Systems.*

larger suspended solids. Figures 16.4, 16.5 and 16.6 show spiral wound UF, hollow fiber MF, and tubular UF systems, respectively.

Microfiltration and UF typically encounter some degree of fouling with suspended solids. Fouling can be particularly severe upon start-up under high transmembrane pressure and flux, which can cause compaction of the initial fouling layer. This results in a dramatic loss of flux and rapid drop in driving pressure, depending of the solids loading. Figure 16.7 compares the flux decline with time for MF and UF membranes. As the figure indicates, MF systems can exhibit higher initial flux, but as suspended solids begin to plug the pores of the membrane (which are larger than the pores in a UF membrane, and hence, more prone to plugging), the flux can decline more rapidly for the MF membrane.

Hollow fiber membrane modules can be backwashed to remove foulants whereas tubular and most spiral configurations cannot be backwashed. Backwashing of traditional spiral-wound modules would break the glue lines holding the membrane leaves together or cause blistering and delamination of the membrane from the backing in both spiral and tubular modules (TriSep Corporation has recently developed a back-washable, spiral-wound module (SpiraSep—US patent 6,755,970), that is used in immersed systems see below).

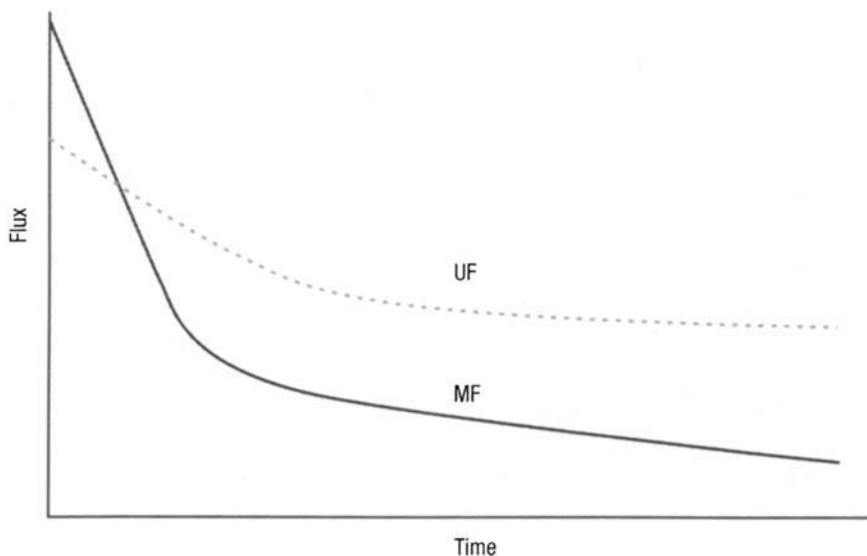


Figure 16.7 Flux decline with time for MF and UF membranes. *Courtesy of Dow Water and Process Solutions.*

Backwashing of HF membranes may need to be conducted frequently, up to 3 or 4 times per hour, depending of the rate of increase in pressure drop. However, in general, a hollow fiber MF or UF system will backwash once every 15 to 90 minutes. Backwashing is enhanced using air scour, where air is bubbled along the membrane surface to promote turbulence and left off solids that have collected on the membrane surface. Air scour, for outside-in applications, may be applied as often as once every backwash. Flow reversal can be used to remove solids blocking the fiber lumens for inside-out feed membranes (air scour can also be used on some inside-out modules such as the Norit Aquaflex 1.5 mm ID and Norit Airlift). Backwashing, along with air scour and flow reversal, minimizes the frequency of chemical cleaning.

A limitation of HF membrane modules is the fact that just as spiral wound modules require pretreatment, so do HF modules, to reduce the size and/or quantity of suspended solids in the feed water, thereby minimizing the chance of plugging the fibers and modules. This is particularly important for inside-out feed HF membranes. Typical pretreatment includes a 150 μ m or smaller screen filter.

Table 16.3 Summarizes the advantages and limitations of tubular, spiral wound and hollow fiber module configurations.

Table 16.3 Advantages and limitation of various module configurations for microfiltration and ultrafiltration membranes.

MF/UF module Configuration	Advantages	Limitations
Tubular	<ul style="list-style-type: none"> • Plug resistant • Easily (mechanically) cleaned 	<ul style="list-style-type: none"> • Large footprint • High capital cost
Spiral Wound	<ul style="list-style-type: none"> • Small footprint • Low capital cost 	<ul style="list-style-type: none"> • Plugging of feed channel • Difficult to clean • Cannot perform integrity tests
Hollow Fiber	<ul style="list-style-type: none"> • Small footprint • Low capital cost • Can be backwashed • Can easily be integrity tested • High membrane area per unit volume (high packing density) 	<ul style="list-style-type: none"> • Plugging of fiber (inside-out feed) • Bridging of fiber bundle (outside-in feed) • Difficult to clean

Microfiltration and UF modules can be operated with a pressurized feed or “immersed” in ambient-pressure solution, with a vacuum (or gravity) on the lumen side of the fiber. Figure 16.8a shows a pressurized MF module, while Figure 16.8b shows a module used in immersed systems. In a pressurized system, water is forced into the membrane under pressure, as shown in Figure 16.9a. Immersed or “submerged” systems rely on suction to pull water through the membrane, as shown in Figure 16.9b. To minimize concentration polarization and fouling of the submerged fibers, two-phase bubbly flow (as illustrated in Figure 16.8b) is used to induce surface shear when operating on water containing a high concentration of suspended solids.

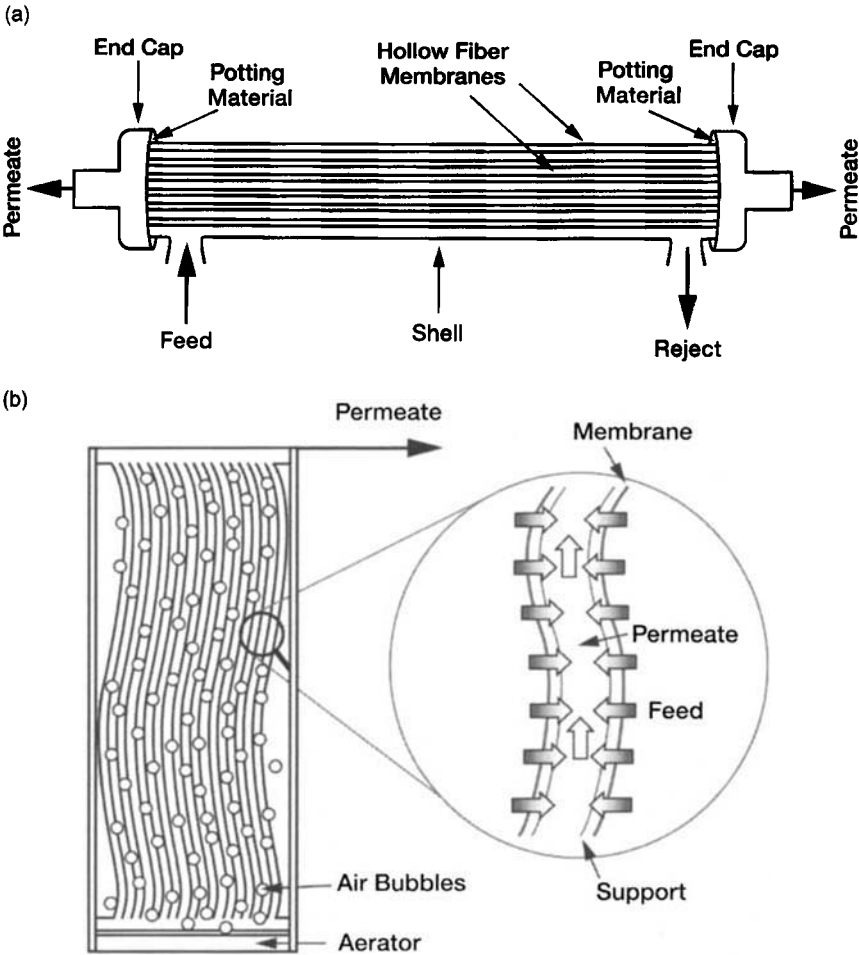


Figure 16.8 Cross-sections of a (a) pressurized MF or UF hollow fiber membrane module and an (b) submerged hollow fiber MF membrane cartridge.

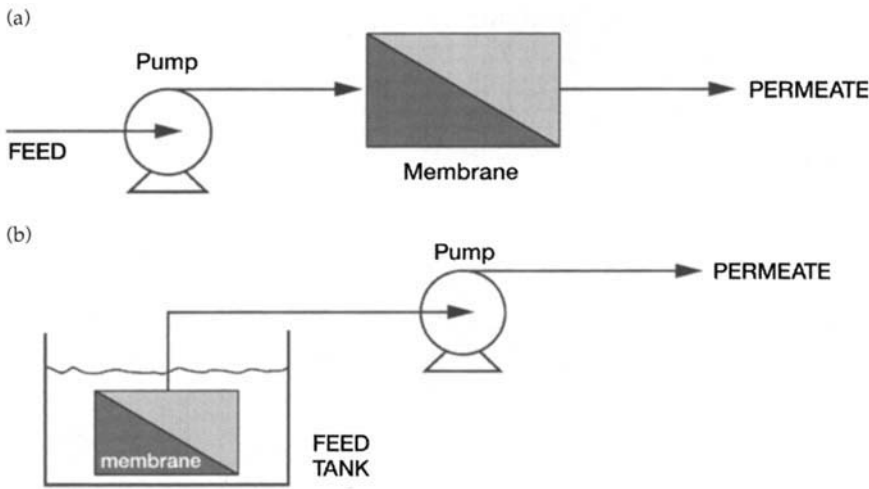


Figure 16.9 (a) pressurized and (b) submerged hollow fibers systems.

Table 16.4 Comparison of pressurized and submerged hollow fiber MF and UF systems.

Parameter	Pressurized Configuration	Submerged Configuration
Operation	<ul style="list-style-type: none">• Dead-end or Cross-flow• Outside-in, Inside-out	<ul style="list-style-type: none">• Dead-end or Cross-flow• Outside-in
Flux	<ul style="list-style-type: none">• Flexible	<ul style="list-style-type: none">• Limited up-turn in flux
Cleaning	<ul style="list-style-type: none">• Less chemical• All chemicals contained in modules & piping	<ul style="list-style-type: none">• More Chemical• Limited to soaking rather than recirculation• Covers and ventilation required over cells
Fouling Control	<ul style="list-style-type: none">• Moderate to Good	<ul style="list-style-type: none">• Moderate to good
Capital Cost	<ul style="list-style-type: none">• Moderate to Low	<ul style="list-style-type: none">• Low
Integrity testing	<ul style="list-style-type: none">• Excellent ability to identify module	<ul style="list-style-type: none">• Identification of module is different and requires removing covers

Table 16.4 shows a comparison of pressurized and submerged hollow fiber MF and UF systems.³

- **Operation:** Pressurized configurations can be operated in either cross-flow or dead-end modes while submerged configuration are essentially only operated in dead-end mode. Submerged systems operate with outside-in flow, while pressurized can be either outside-in or inside out flow.
- **Flux:** Both pressurized and submerged configurations can be operated below design flux without difficulty. If operation above design flux is required for short periods of time, pressurized systems are preferred, as they operate under a positive trans-membrane pressure; positive pressure can always be increased, while vacuum is limited.
- **Cleaning:** Pressurized configurations required less chemical during cleaning than submerged systems. This is because the entire tank containing the submerged membrane cartridge must be filled with chemical to clean the membranes (note that spacers are used to occupy space to minimize this factor).
- **Fouling Control:** Pressurized and submerged hollow fiber configurations with out-side-in flow both suffer from blockage of the fiber bundles that tends to be self-accelerating. Hence, the need for two-phase bubbly flow to minimize accumulation of foulants among the fibers in a submerged system to mitigate this issue.
- **Capital Cost:** Submerged configurations have a slight cost advantage over pressurized configurations when only the equipment supplied by the membrane supplier is considered. Submerged systems lack pressure vessels and the plumbing is simpler, but tankage must be supplied by the constructor. Additionally, submerged system will need covers over the cells and ventilation systems to avoid chemical fuming and other issues that are not a concern with pressurized systems.

Hollow fibers are generally used for submerged applications, although flat sheets, and more recently, spiral wound modules have come into use in submerged systems. In general, pressurized and submerged systems compare favorably, with the submerged

configuration enjoying a slight cost advantage, particularly for larger systems. Both configurations are used as pretreatment to RO with success.

It is important to take into account permanent fouling of the MF or UF membranes by over designing the system. Despite the dense surface of these membranes, pore fouling can occur, particularly at higher membrane driving pressures (typically caused when the system is operated at high permeate flux or high trans-membrane pressure (TMP)). Under constant pressure, MF and UF membranes can lose 50% – 75% of water flux between backwashes. Pilot testing is necessary to determine the degree of permanent fouling as well as the backwash and air scour frequencies and the chemical programs that are required. Pilot testing is particularly important in colder climates where the seasonal range for temperature is great, as temperature affects the flux through the membranes (higher temperature leads to higher flux and, conversely, lower temperature leads to lower flux in a manner similar to RO membranes).

As discussed in Chapter 8.1.9, MF and UF membranes can delay the onset of microbial fouling of RO membranes, but by themselves are not fully effective. Nutrients, in the form of low-molecular weight organics, can pass through these membranes such that any post introduction of microbes into the RO feed water will result in microbial fouling. Therefore, the use of chlorine is recommended in conjunction with these membrane processes to minimize the potential for microbial fouling of RO membranes.

16.1.1 Microfiltration

Microfiltration is a pressure-driven membrane separation technology used to separate particles from solution (see Figure 1.1). This technology uses microporous membranes with pore sizes ranging from about 0.1 up to about 3 microns. However, due to particle adsorption onto the surface of the pores, and the collection of particles on the top of the membrane (both of which serve to block pores), MF membranes can often remove particles smaller than the rated pore size.⁴ Microfiltration cannot, however, remove all colloidal material because colloids can be as small as 0.01 microns in diameter.⁵ Since MF membranes do not reject ions, osmotic pressure is not a concern as it is with RO systems. Typical operating pressures for MF systems range from about 10 – 100 psi with 30 – 40 psi being typical. Industrial flux rates are up to 100 gfd, depending on the nature of the feed water.

Applications of MF include:

- Water treatment—removal of suspended solids, including up to 4-log removal of bacteria (a major application growth area¹),
- Wastewater treatment—biomass sedimentation using membrane bioreactors (MBRs), tertiary treatment of municipal waste for reuse,
- Food & beverage treatment—clarification of liquids (fruit juices, beer),
- Resource recovery—recovery of inks, textile sizing agents, and electrodeposition paints.

Microfiltration as pretreatment to RO is used for bulk removal of suspended solids and bacteria. Microfiltration is a barrier technology that can in many cases replace conventional clarification and filtration. Advantages of MF over conventional clarification and filtration include the following:

- Minimal need for treatment chemicals—coagulants, flocculants, acid or caustic (coagulants are used when organic removal is required),
- Consistent effluent quality, regardless of raw feed water quality,
- Compact system,
- Simple automation—minimal operator attention required.

See references 1, and 6 – 8 for more detailed discussions about microfiltration.

16.1.2 Ultrafiltration

Ultrafiltration is a pressure-driven membrane separation technology used to separate particles and high-molecular-weight soluble species from solution. Ultrafiltration also uses microporous membranes with a pore size distribution of 0.005 to 0.1 microns, an order of magnitude smaller than the pore sizes for MF (see Figure 1.1). In general, because of the smaller pores, the operating pressure required for transport through a UF membrane is greater than that for an MF membrane, generally 5 – 150 psi with 30 – 75 psi being typical (some UF membranes, such as the Norit PES membrane,

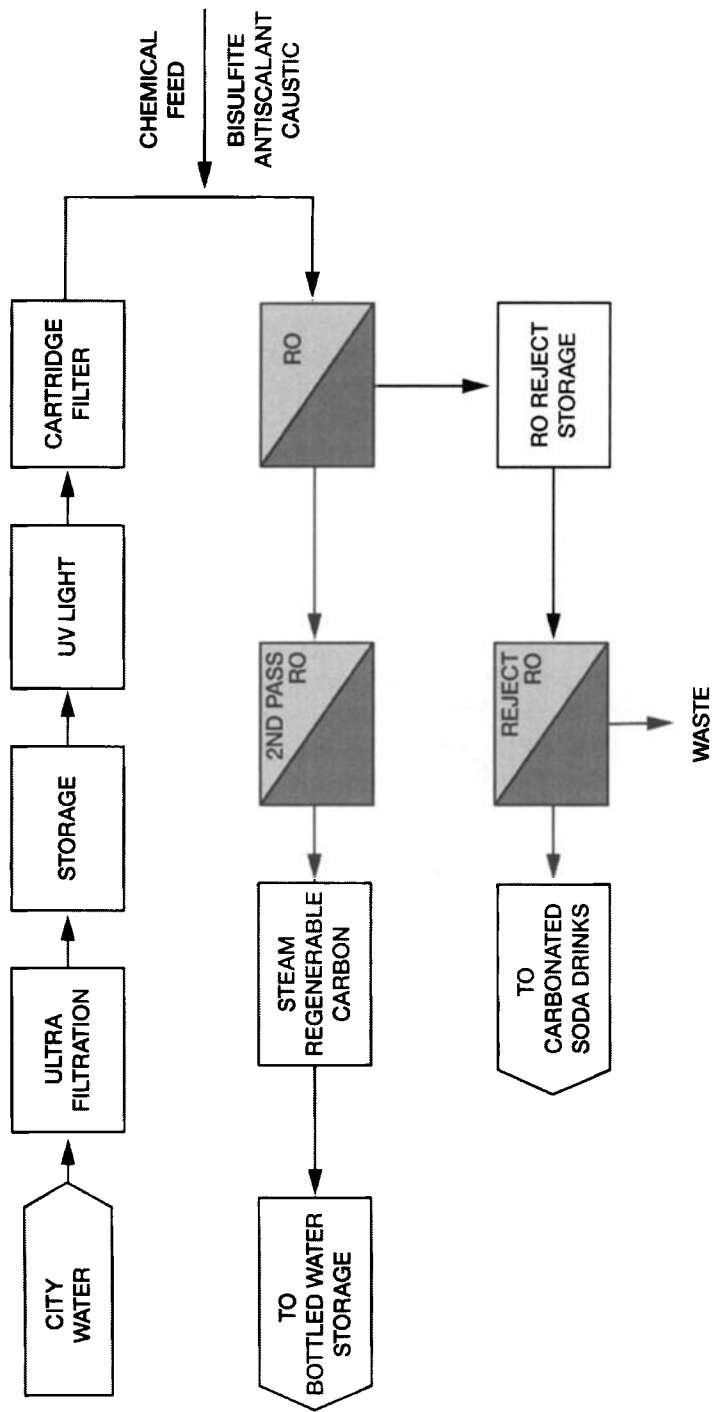


Figure 16.10 Typical water treatment process flow diagram for a beverage facility producing both bottled water and carbonated soda drinks.

require a very low net driving force of 1.5 – 2.0 psi on clean membranes assuming proper flux selection). Typical industrial flux rates for UF range from 5 to 70 gfd, depending on the nature of the feed water.

As discussed above, UF membranes have smaller pores than MF membranes indicating that their rejection of suspended solids and bacteria are greater or “tighter” and that for MF. For example, the California Department of Public Health gives UF membranes a 4-log removal credit for viruses, while MF typically receives only a 0.5 log-removal credit for viruses. Species that are retained by a UF membrane are typically defined as those whose molecular weight is greater than the molecular weight cut off (MWCO) of the UF membrane. The MWCO of a UF membrane describes the ability of that membrane to retain 90% of a challenge macrosolute (such as glycol, dextran, or protein)⁹. There is no international standard for MWCO, so that membranes from different manufacturers cannot be directly compared on the basis of MWCO only.

Applications of UF include:

- Food & beverage—recovery of proteins for milk and whey, sugar concentration of fruit juice, removal of bacteria and suspended solids and organics from bottled drinking water,
- Wastewater—removal of free oil,
- Municipal water—removal of bacteria and large viruses from potable water and wastewater for reuse.
- RO pretreatment—reduction in concentration of suspended solids and high molecular-weight organics.

The application of UF for RO pretreatment is growing, as users are interested in reducing the size of their treatment facilities and reducing or eliminating chemical treatments. Although capital cost is an issue, UF can be cost effective when used to remove bacteria and dissolved high-molecular weight organics from RO feed water and as a polisher for reduction of suspended solids and silt density index. Ultrafiltration in the tubular configuration can also be used for bulk removal of suspended solids prior to RO. Ultrafiltration is a barrier technology providing consistent effluent quality including 6-log bacterial removal and SDIs typically less than 2 (turbidity less than 0.02 NTU). Ultrafiltration prior to RO is commonly seen in the production of bottled drinking water, even when potable water is used as

Table 16.5 Rejection of ions exhibited by NF and RO membranes.

Species	Nanofiltration Rejection (%)	Reverse Osmosis Rejection* (%)
Calcium	75 – 98	93 – 99
Magnesium	70 – 98	93 – 98
Sodium	45 – 95	92 – 98
Chloride	1 – 95	92 – 98
Sulfate	95 – 99	96 – 99
Bicarbonate	40 – 95	85 – 95
Fluoride	25 – 95	92 – 95
Silica	5 – 95	90 – 98
Total Dissolved Solids	65 – 95	90 – 98

*Polyamide composite membranes

operates at lower pressure than RO, typically 50–150 psi. Nanofiltration falls between RO and UF on the filtration spectrum shown in Figure 1.1. Table 16.5 compares the general differences in rejection of species between NF and RO membranes.

As shown in the table, NF membranes offer a wide selection of rejection capabilities. In general, rejection of divalent ions is greater than that for mono-valent ions for NF membranes, particularly for the “loosest” of NF membranes. Reverse osmosis membranes, on the other hand, exhibit high rejection of both mono-valent and divalent ions (although rejection of divalent ions is also greater than that of mono-valent ions—see Chapter 4.2.2). The addition of trace additives and the different dissociation constants of the piperazine found in the FilmTec NF membrane (Figure 16.11) is used to yield the wide range of solute transport through this membrane.

In simple terms, nanofiltration membranes reject species based on size or charge of the particle, depending on the charge of the membrane.^{1, 11, 12} For example, cationic NF membranes have negatively-charged groups attached to the polymer backbone and consequently, sulfate, which is negatively charged, is rejected to a greater extent by an cationic NF membranes than calcium, which has a larger hydrated radius but is positively charged.^{1, 11} This is a result of the negative charge on the membrane repulsing the negatively-charged

sulfate ion (Donnan exclusion effect). The same principle applies to anionic membranes and cationic species.

Fouling and scaling mechanisms are similar for spiral-wound NF and RO membranes. In general, NF feed water should meet the following characteristics to prevent fouling with suspended solids (refer to Table 7.1 for a more detailed description of spiral-wound RO feed water requirements):

- SDI < 5 (< 3 is preferred)
- Turbidity < 1 NTU (< 0.5 NTU is preferred)

Hence, traditional spiral wound NF membranes require the same level of pretreatment as spiral-wound RO membranes, as well as the same flux and flow rate considerations with respect to feed water quality (see Chapters 9.9 and 9.4).

New generation NF membranes have been developed that exhibit lower fouling tendencies, making them applicable to high organics and high-color applications.¹¹ While conventional RO and NF membranes have a rough surface and a negative or positive surface charge, these new NF membranes have a smooth surface and more neutral surface charge, making them more resistant to organic fouling. These low-fouling NF membranes can operate with TOC concentrations ranging from 10 – 20 ppm and color up to 90 APHA units.¹⁰ Examples of these new NF membranes include the Hydranautics' ESNA1-LF, LF2, and LF3.

Applications of NF include:

- Water treatment—water softening and color removal for potable water applications
- Wastewater treatment—color removal from pulp and paper wastewater
- Resource recovery—recovery of valuable, lower-molecular weight products in the drug, semiconductor, textile, metal-plating, and food industries.⁹

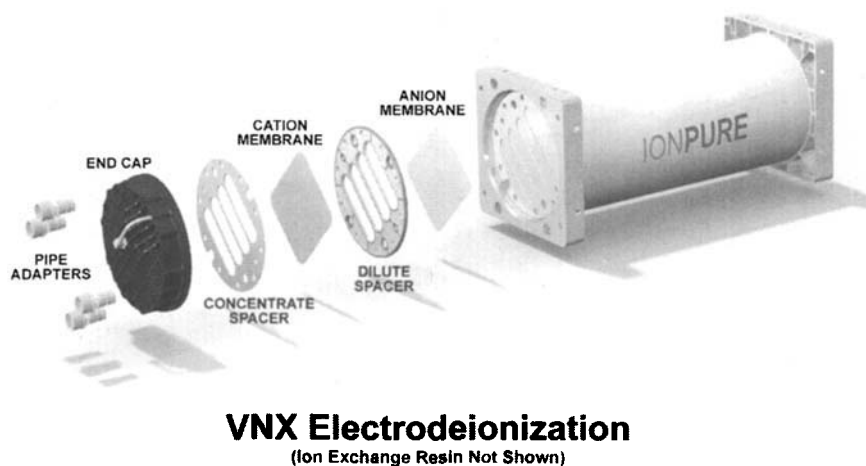
For RO pretreatment, NF is typically used to pre-soften and reduce color from RO feed water (when appropriate NF membranes are employed). Nanofiltration replaces sodium softening (for hardness removal) and augments clarification (for color removal).

16.3 Continuous Electrodeionization

Continuous electrodeionization (CEDI—the continuous process subset of electrodeionization (EDI) that is sometimes referred to as

just continuous deionization or CDI® (CDI is a registered trademark of Siemens Water Technologies)) is an electrically-driven membrane separation technique under development since the mid-1950's and successfully commercialized in the late 1980's.¹³ Continuous electrodionization is used to remove trace dissolved solids from water. This technology relies on a D.C. current that is applied to cation- and anion-exchange membranes with ion exchange resin arranged in a "stack" configuration to effect the separation of solutes from water.

Figure 16.12 shows the expanded view of a conventional, stacked-disk plate-and-frame CEDI module (stack). Inside one end of the module is the cathode and inside the other end is the anode. Between the cathode and anode are layers of flat-sheet cation- and anion-exchange membranes separated by spacers that act as alternating diluting (product) and concentrating (reject) compartments. Resin is placed in the diluting compartments, as shown in Figure 16.13a. Note that in the newer "all filled" configurations, all compartments are filled with ion exchange resin as shown in Figure 16.13b. An electrical potential is applied to the stack (100 – 600 volts DC at 3 – 10 amps) that drives the cations toward the cathode and anions toward the anode. (The amount of current drawn from the power supply is proportional to the concentration of ions in the solution. Typically, 4 ppm feed to a CEDI system will draw about 3 amps.¹⁴) Cations pass through the cation-exchange membranes but not through the anion-exchange membranes, while anions pass through the anion-exchange membranes but not through the cation-exchange membranes. (Note



VNX Electrodeionization
(Ion Exchange Resin Not Shown)

Figure 16.12 CEDI stacked-disk, plate-and-frame configuration. *Courtesy of Siemens Water Technologies—Ionpure Products.*

that the ion exchange membranes are not permeable to water; only ions are transported through them.) In this manner, every other compartment becomes dilute in ions while the other compartments become more concentrated in ions. The ion exchange resin facilitates the transfer of ions in low ionic-strength solutions.

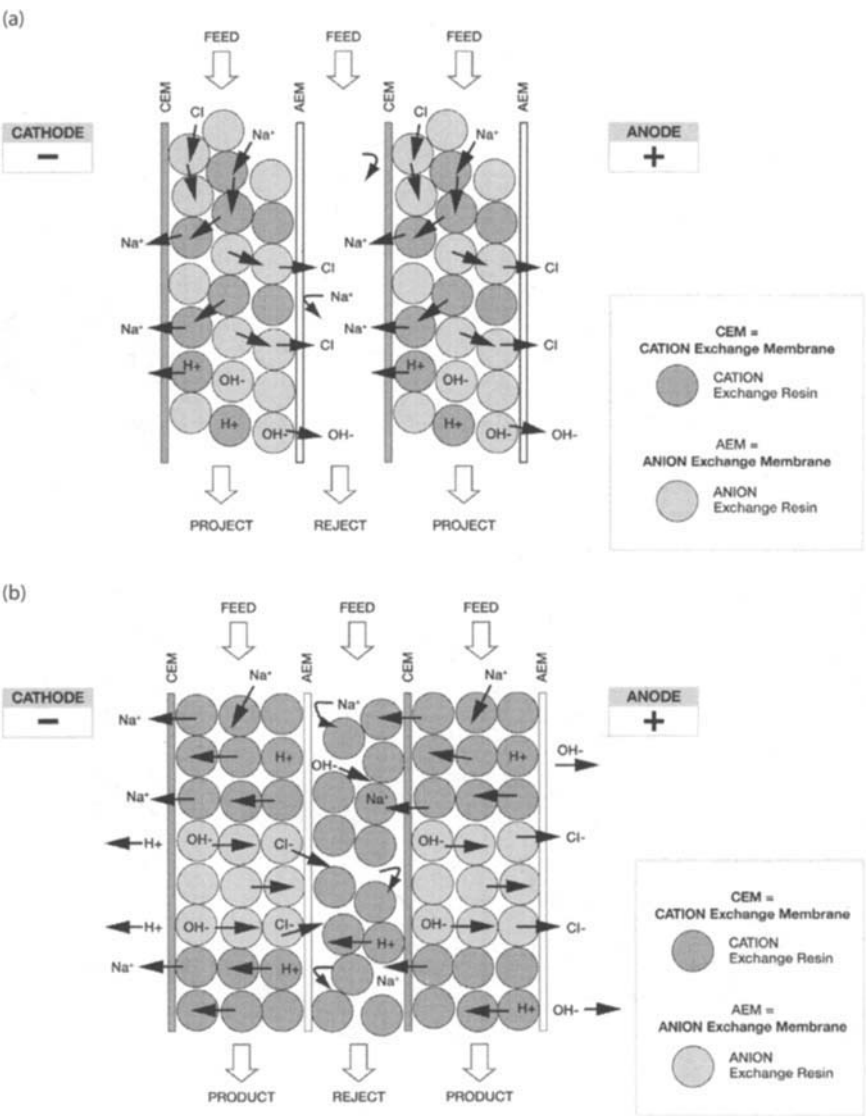


Figure 16.13 Transport through cation- and anion-exchange membranes in a) "unfilled" and b) "all filled" CEDI stacks. *Courtesy of Siemens Water Technologies—Ionpure Products.*

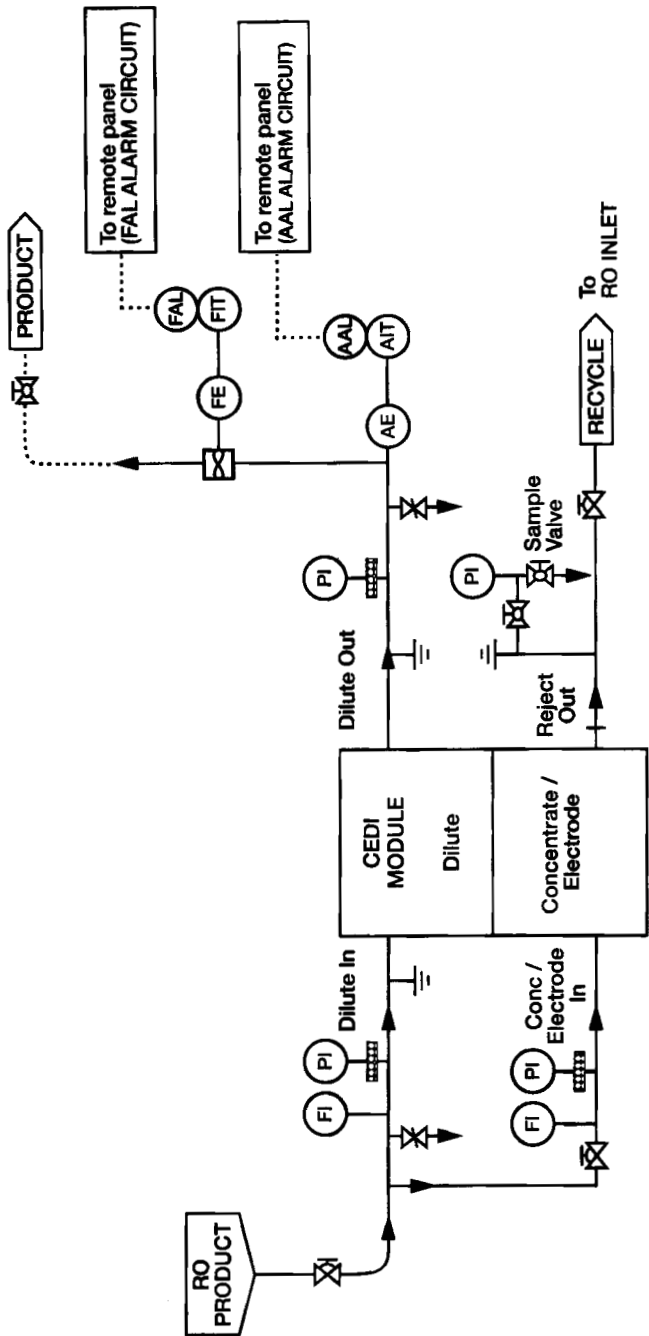


Figure 16.14 Typical process flow diagram for an all-filled CEDI configuration. Courtesy of Siemens Water Technologies—Ionpure Products.

Figure 16.14 shows a typical process flow diagram for an all-filled configuration, while Figure 16.15 shows a typical process flow diagram for an unfilled configuration with concentrate recycle. Recycle is used to achieve higher recovery while maintaining good velocity through the concentrating compartments. The concentrate recycle also increases the concentration in the concentrated compartment, thereby reducing the resistance of the stack to the DC current. In some cases, brine injection is used to supplement the concentration recycle, particularly where the ionic concentration of the feed water is very low. The all-filled configuration does not need brine recycle or brine injection. This results in lower capital cost for the all-filled configuration, as the brine recirculation and brine metering pumps and associated piping are eliminated. All-filled configurations offer less electrical resistance and suffer less salt bridging than the unfilled configuration (see discussion on salt bridging below).

Equations 16.1 and 16.2 show the reactions at the cathode and anode, respectively. These equations indicate that hydrogen and oxygen gases are produced at the electrodes. Typically, 7.5 ml/min of hydrogen and 2.7 ml/min of oxygen are produced at 25°C and 14.7 psig. Equation 16.3 indicates that chlorine gas may also be generated at the anode. Concentrations of chlorine gas range from non-detectable up to 8 ppm, depending on the configuration of the CEDI module and whether or brine injection or concentrate recycle is employed.¹⁵ Chlorine gas is more likely to be generated when more chloride is present in the concentrate compartment, as is the case when brine injection is used. These gases require removal that is usually accomplished using water passing over the electrodes and then venting the gases from this water flush stream.¹⁵



Some companies market spiral wound CEDI modules, such as the DOW™ EDI spiral wound module (formerly Omexell—DOW is a trademark of Dow Chemical Company, Inc., Midland, Michigan). Figure 16.16 shows the cross section of the Dow module.¹⁶ As the figure shows, RO permeate enters the module on one end, while CEDI effluent exits the other end; concentrate is sent spiraling into the metal center tube of the module and exits out of this tube. A titanium

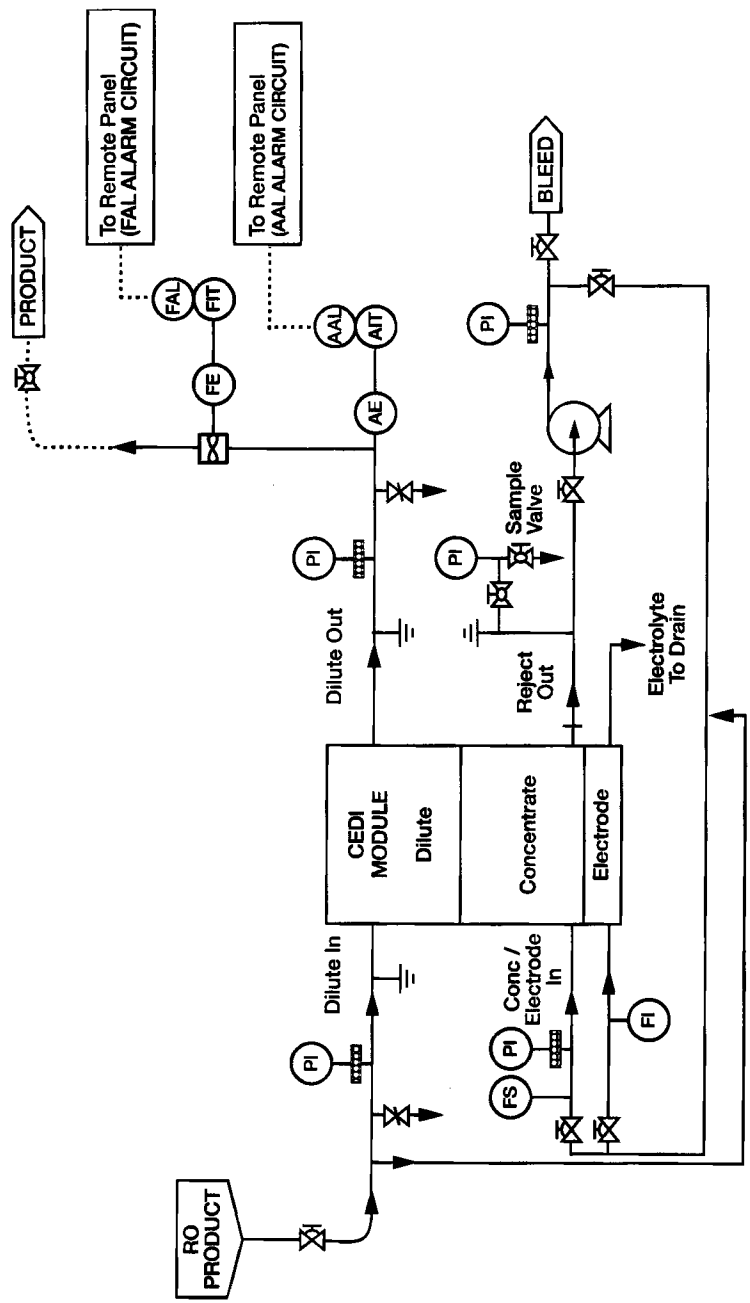


Figure 16.15 Typical process flow diagram for a CEDI configuration with unfilled concentrate spacers. Courtesy of Siemens Water Technologies—Impure Products.

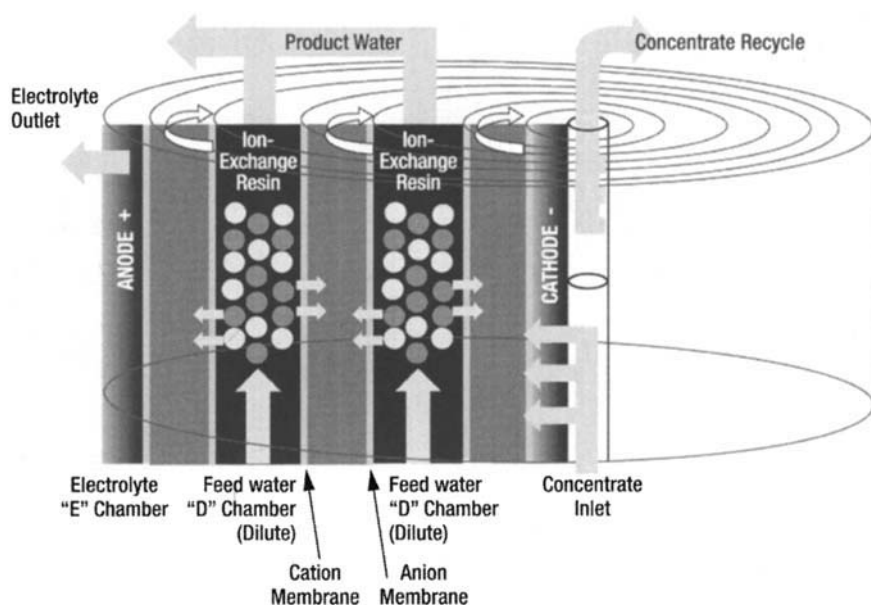


Figure 16.16 Cross-section of a DOW™ EDI—CEDI spiral wound module.
Courtesy of Dow Water and Process Solutions.

cylinder is placed inside the fiberglass outer shell to act as the anode. The cathode is the metal center tube. In the case of this spiral module, only the diluting compartments are filled with ion exchange resin. Hence, a recycle of the concentrate solution is required. Figure 16.17 shows a process flow diagram of a spiral-wound CEDI module.

The major advantage of the spiral configuration over a plate-and-frame configuration is that there is minimal leakage associated with the spiral configuration. The spiral wound module does not require periodic tightening of nuts and bolts to prevent leaks, unlike plate-and-frame modules. Limitations of the spiral configuration include inferior current and flow distribution relative to plate-and-frame modules, as well as difficulty in assembly and field membrane replacement.¹⁷

A CEDI system can produce up to 18-megohm-cm water at 90-95% water recovery. Recovery by the CEDI system is a function of the total hardness in the feed water to the system. In general, 95% recovery can be realized at a feed water hardness of less than 0.1 ppm as calcium carbonate.¹⁶ This is typically attained if the pretreatment to the CEDI consists of either 2-pass RO or sodium-cycle softening followed by RO. Recovery that is achievable is a function of the feed

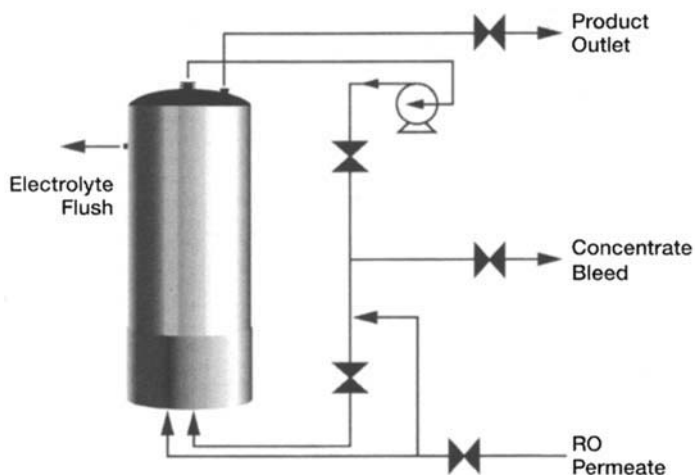


Figure 16.17 Process flow diagram of a spiral wound CEDI module. *Courtesy of Dow Water and Process Solutions.*

Table 16.6 Continuous electrodeionization recovery as a function of feed water hardness for an E-Cell (GE) module.¹⁸

Feed Water Hardness (ppm as CaCO_3)	CEDI Recovery (%)
< 0.10	95
0.10 – 0.50	90
0.50 – 0.75	85
0.75 – 1.00	80

water hardness concentration and varies with manufacturer as well. Table 16.6 shows how recovery is a function of feed water hardness for a GE—E-Cell® module (E-Cell is a registered trademark of General Electric Company, Fairfield, Connecticut).¹⁸

Continuous electrodeionization systems can achieve 95% rejection of boron and silica, and 99+% rejection of sodium and chloride. This performance is possible due to voltage-induced dissociation of water that effectively regenerates a portion of the resin thereby allowing removal of weakly ionized species such as silica and boron.¹⁹ In fact, the boron in the effluent from a CEDI system can be lower than that in the effluent from a mixed-bed ion exchange system.¹³

High feed water carbon dioxide is the leading cause of poor product quality in the effluent from a CEDI unit. This is because carbon dioxide is converted to carbonate and bicarbonate within the stack, and thus it contributes to total exchangeable anions. This must be taken into account when designing the system. Carbon dioxide concentrations below 5 ppm can reduce the removal of ions from the feed water, particularly weakly ionized species such as silica and boron.¹³ In an effort to account for the carbon dioxide and its draw on the DC current, a “conversion factor” of 2.79 micro-Siemens per centimeter for every 1 ppm carbon dioxide is added to the measured conductivity of the feed water to the CEDI unit.¹⁹ Carbon dioxide can be removed prior to the CEDI system using a membrane degasification system such as a Membrana Liqui-Cel® membrane contactor (Liqui-Cel is a registered trademark of Celgard LLC, Charlotte, NC) thereby minimizing the power required by the CEDI system and improving the removal efficiency of other ions.

Note that silica, is weakly ionized and will also contribute to the draw on DC current. A conversion-factor of 1.94 micro-Siemens per centimeter for every 1 ppm silica is added to the measured conductivity of the feed water to account for this. Although the conversion-factor is greater for silica than for carbon dioxide, silica has much lesser effect on the DC current draw than carbon dioxide because of the relatively low concentration of silica in CEDI feed water as compared to carbondioxide.

A significant amount of pretreatment is required to minimize fouling and scaling of the membranes in a CEDI system. Table 16.7 lists general feed water quality requirements for CEDI systems.^{15, 21} Due to the stringent feed water quality requirements, most CEDI systems are preceded by RO. Common configurations used to pre-treat CEDI feed water include the following:¹⁵

- Antiscalant → RO
- Softening → RO
- Antiscalant → RO → softening
- Antiscalant → RO → RO

Because of the high-quality feed water sent to a CEDI system, the concentrate from the CEDI is very low in dissolved solids and is often recycled to the influent of the RO system. This reduces the overall waste generated by the system and increases the water quality to the RO through dilution of the source water.

Table 16.7 General feed water quality requirements for CEDI systems. Adapted from specification sheets for various manufacturers.

Constituent	Typical Feed Quality	Units
Total Exchangeable Anions (TEA)	< 25	ppm
Total Hardness	< 0.5*	ppm CaCO ₃
Iron, Manganese, Hydrogen Sulfide	< 0.01	ppm
Silica (soluble)	< 0.5 (some up to 1.0)	ppm
Conductivity	< 65	micro-S/cm
pH	5 – 9 (range 4 – 11)	
Free Chlorine	< 0.05	ppm
Other Oxidizing agents	ND**	ppm
Carbon Dioxide	< 10	ppm
Total Organic Carbon (TOC)	< 0.5	ppm
Oil & Grease	ND**	ppm
Color	< 5	APHA
SDI	< 1.0	
Turbidity	< 1.0	NTU
Temperature	35 – 113	°F
Pressure	25 – 100	psig

*See Table 17.4

**None detectable

Despite efforts to comply with the limitations on feed water quality, CEDI systems can still foul and scale with microbes, organics, iron and manganese, and calcium- and silica-based scales. This usually occurs due to upsets in the pretreatment system or a deficiency in the system design that result in excursion in feed water quality to the CEDI system. Organics, metals, hardness, and silica problems are usually found on the membranes and sometimes on the resin (as is the case with organics). Biofouling is typically found on the

concentrate spacers outside of the electric field.¹⁵ System with recirculating concentrate experience greater biofouling issues than the once-through (all-filled) systems.¹⁵

If the membranes do foul or scale, they can often be cleaned. A typical cleaning frequency is once per year. Cleaners include

- *Sodium chloride/sodium hydroxide mixture*: used for organic fouling in a manner similar to a “brine squeeze” of anion resin
- *Hydantoin (Halane)*: used for microbial fouling—should be used infrequently
- *Peracetic acid*: used as a sanitizing agent to prevent microbial fouling
- *Sodium carbonate/hydrogen peroxide mixture*: used for removal of biofilm
- *Hydrochloric acid*: used for scale
- *Sodium hydroxide*: used for microbial fouling and silica scale

Consult with the vendor for details on how to properly clean specific CEDI systems. Most CEDI systems can handle temperatures up to 45°C during cleaning (see Table 16.7). Some CEDI units may be heat sanitized at up to 80°C.

Continuous electrodeionization systems can also suffer mechanical failure, including salt bridging and electrical arcing. Salt bridging occurs when leaks, such as water wicking out of membrane edges that are left exposed to the environment, evaporate and leave behind salts. The amount of salt can build up until bridging occurs. This provides another path for the DC current, leading to arcing and module damage.¹⁵

The system can also degrade due to exposure to free chlorine. Note that this problem is usually encountered when chloramines are feed to the RO. In such cases, the amount of free chlorine in the RO product can be greater than that in the feed. Hence, total chlorine must be removed prior to the RO membranes before it even gets to the CEDI system.

Continuous electrodeionization is primarily used as an alternative to ion exchange. Because of the extensive pretreatment required by CEDI systems, the technology has grown into a polisher for RO (see Figure 16.18). Continuous deionization can achieve mixed-bed water quality of RO permeate without the need to store and

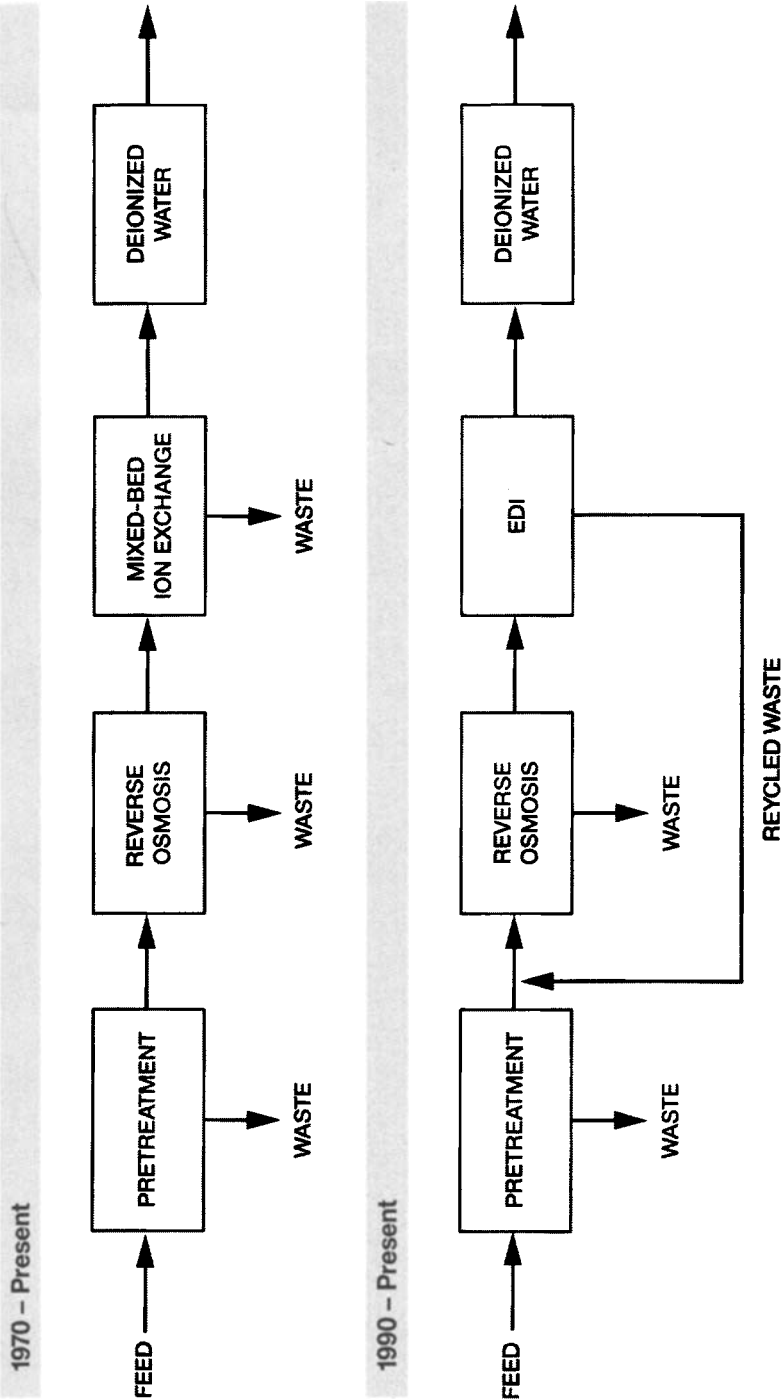


Figure 16.18 Evolution of membrane deionization.

handle acid, caustic, and regeneration waste associated with conventional ion exchange. As a result, CEDI systems take up less space than traditional ion exchange that is equipped with regeneration equipment. Costs for CEDI systems have also decreased relative to mixed-bed ion exchange, as shown in Figure 16.19.

The pharmaceutical industry lead the way in adoption of CEDI for the production of ultrapure water. Since the early 1990's, the power industry has been employing CEDI as a polisher for RO effluent for steam generation. Other industries currently using CEDI include general industry for boiler make-up or high-purity process applications, including semiconductor manufacture. Commercially-available industrial CEDI modules range in size from less than 1 gpm to 80 gpm.

Manufacturers of CEDI modules include:

- Siemens Water Technologies (Ionpure®)
- GE (E-Cell®)
- Dow Water and Process Solutions (Omexcell)
- Snowpure (Electropure™)
- Christ Water Technology Group (Septron®)
- Millipore (Elix®)

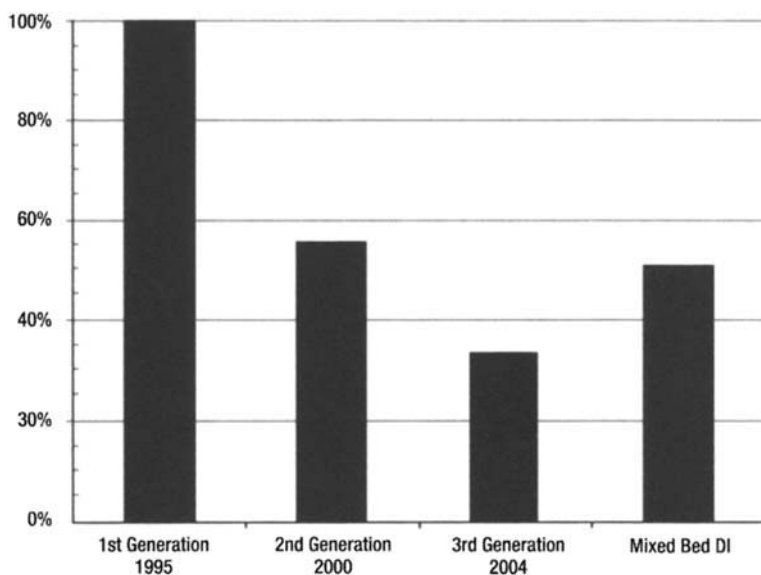


Figure 16.19 Costs of CEDI systems relative to mixed-bed ion exchange systems that follow improvements to CEDI technology. Assumes 120 gpm system. *Courtesy of Siemens Water Technologies—Ionpure Products.*

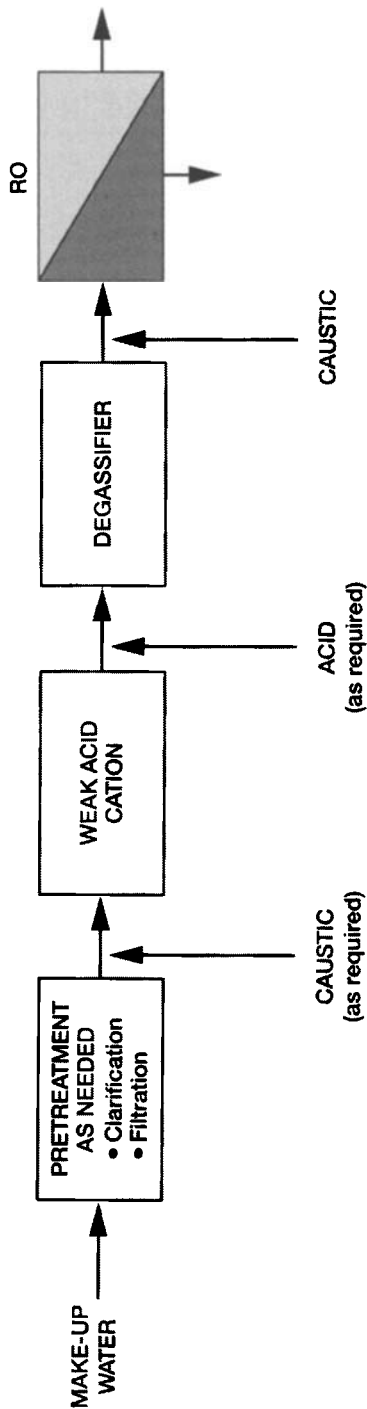


Figure 16.20 Process flow diagram of a typical HERO™ process.

16.4 HERO™ Process

High efficiency reverse osmosis (HERO) is a patented process (US Patent # 5925255, Debasish Mukhopadhyay, 1999) originally developed to treat high silica water for the microelectronics industry. Its use has expanded to power and zero liquid discharge applications (see Chapter 15.4.3). Features of the process include:

- High water recovery, typically 90 – 95%
- High rejections of species, including weakly ionized compounds such as boron and fluoride
- Significant reduction in membrane fouling from organics and microbes

The HERO process includes the following process steps:

- Hardness/bicarbonate alkalinity removal
- Dissolved carbon dioxide gas removal
- Caustic addition to pH 10 – 10.5

Figure 16.20 shows a simplified process flow diagram of a typical HERO process. The preferred method of hardness and bicarbonate alkalinity removal is simultaneous removal in a weak acid cation (WAC) ion exchange unit. Caustic is sometimes added prior to the WAC unit to improve the efficiency of the hardness/alkalinity removal process. Hardness removal via the WAC unit enables the RO system to operate at high recovery without fear of hardness scaling. The exchange of hardness for hydrogen ions in the WAC unit decreases the pH of the water, converting much of the alkalinity to carbonic acid and carbon dioxide. Additional acid is sometimes added after the WAC unit to complete the bicarbonate alkalinity conversion to carbon dioxide gas. Carbon dioxide is then removed in the degassifier (either conventional or membrane-based). Caustic is added after the degassifier to raise the pH to about 10 – 10.5 prior to the RO. Raising the pH to this high level does several things:

- Increases the solubility of silica, thereby reducing its tendency to scale. Solubility of silica at ambient temperature and pH 11 (typical reject pH in a HERO system) is greater than 1,500 ppm as compared to a solubility of about 120 – 160 ppm at pH 8 (see Chapter 7.8

and Figure 7.2, which shows how the silica solubility increases exponentially at concentrate pH greater than about 8).

- Increases the ionization of silica, thereby increasing its rejection by the RO membrane.
- Increases the ionization of weakly ionized species such as boron, fluoride, and organics as TOC, thereby increasing their rejection by the RO membrane.
- Destroys or disables the ability of biological organisms to propagate, thereby eliminating microbial growth on the RO membranes.
- Ionizes organics to organic acids, thereby increasing their rejection by the membrane and decreasing their tendency to foul the membrane.

In many cases, additional pretreatment is required prior to the HERO process. Additional pretreatment can include cold lime softening, sodium softening, and ultrafiltration or microfiltration. These pretreatment unit operations are required when the total hardness is very high and/or the concentration of suspended solids is high.

High efficiency RO is not practical for every-day RO applications due to the relative complexity and high capital and operating cost involved. However, for difficult feed streams that require high recovery, HERO is well suited. Advantages of HERO over conventional RO for these applications include:

- Very high recovery, including high silica feed water
- Less space required due to higher water flux and smaller RO (HERO systems typically operate at 25 – 30 gfd)²²
- Reduced membrane cleaning due to the elimination of bacterial growth on the membranes and reduced silica scaling
- Reduced capital cost for larger (greater than about 50 gpm) systems due to smaller RO (higher flux rates)²³
- Lower operating costs due to lower operating energy and fewer specialty chemicals (e.g., antiscalants)

Limitations of HERO for ZLD include the number of unit operations required for treatment and the need for significant chemical application and sludge disposal. Other limitations with the HERO

process revolve around the weak acid cation (WAC) unit that is used to remove hardness associated with alkalinity.

- The pH in the effluent from the WAC unit is typically about 4.5. At this pH, the organics in solution drop out of solution and it is difficult to re-solublize them when the pH is raised. This leads to higher-than expected organic fouling of the RO membranes.
- In many applications, single-stage solids contact clarifiers are used for bulk removal of hardness prior to the HERO process (via lime softening). For applications where cooling tower blowdown is being treated, the single stage softening does not allow sufficient time to break up dispersants used in the tower. Hence, the effluent from the unit is significantly higher than expected. This chelated hardness leaks through the WAC unit, and the RO experiences hardness scaling, particularly at the high recoveries and fluxes often used with HERO.

High efficiency RO is often used in zero liquid discharge (ZLD) applications. The HERO process is used prior to the thermal equipment to reduce the size and energy required by the thermal system. In many cases, the brine concentrator can be eliminated entirely, such that the concentrate from the HERO process feeds directly to the crystallizer.

Suppliers of the HERO process under license include GE and AquaTech.

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FREQUENTLY ASKED QUESTIONS

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Frequently Asked Questions

This chapter answers some common questions about RO in general as well as operational and equipment-related questions.

17.1 General

17.1.1 What Is Reverse Osmosis Used For?

An RO system is designed to remove ions from solution. Rejection of most ions ranges from about 96% to 99+%, depending on the nature of the ion and the type of membrane used (see Chapter 4.2).

Although RO membranes also act as barriers to suspended solids, it is not recommended that they be used for this purpose. The membranes will foul with suspended solids, resulting in higher operating pressure, frequent membrane cleaning, and shorter membrane life. To avoid fouling, pretreatment is required to remove suspended solids from the RO feed water (see Chapter 8).

The most common uses of RO are for desalination of seawater and brackish water for potable and industrial applications. However, as demand for fresh water grows, RO is being pressed into service for wastewater and reuse applications. These will require extensive pretreatment, sometimes involving other membrane technologies such as micro- or ultrafiltration, to minimize fouling of the RO membranes (see Chapter 16).

17.1.2 What is the Difference Between Nanofiltration and Reverse Osmosis?

Nanofiltration (NF) and RO are closely related in that both share the same composite membrane structure and are generally used to remove ions from solution. However, NF membranes use both size and charge of the ion to remove it from solution whereas RO membranes rely only on “solution-diffusion” transport to affect a separation (see Chapters 16.2 and 4.1, respectively). Nanofiltration membranes have pore sizes ranging from about 0.001 to 0.01 microns, and therefore,

the rejection of ions in solution by an NF membrane is not as good as that by an RO membrane (see Figure 1.1). Because NF is similar to RO, but with lower rejection, NF is sometimes called “loose” or “leaky” RO.¹ Nanofiltration is commonly used to “soften” potable water or to remove color and organics from RO feed water.

17.1.3 What is Data Normalization?

Data normalization is a method used to understand the performance of the membranes in an RO system (see Chapters 11.3 and 12). Performance, namely permeate flux, salt rejection, and pressure drop, are all functions of operating conditions, such as temperature and pressure, and functions of the degree of membrane fouling, scaling, and degradation. Since these conditions are always changing, comparison of actual data is difficult, as there are no common reference conditions. Normalizing data takes out the changes in operating conditions (temperature, pressure, and concentration), such that the only changes in normalized performance are due to membrane fouling, scaling, and degradation. Normalization provides a common reference point, start-up conditions, that all data is compared to. Hence, normalizing data allows the user to determine the condition of the membranes, be they fouled, scaled, or degraded.

17.1.4 How do SDI and Turbidity Correlate?

Silt density index (SDI) and turbidity are only loosely related. In general, the higher the feed water turbidity, the higher the SDI. However, the converse is not always true. Low turbidity, less than 1 NTU, can still correspond to high (greater than 5) SDI. This is particularly important to consider when using potable water as the feed source, especially if the ultimate source is surface water. City water generally has a turbidity less than 1 NTU, but will often have SDI greater than 5. Hence, it is not uncommon to install multimedia filters as RO pretreatment on city water sources.

17.1.5 Why Does the pH Drop from the RO Feed to the RO Permeate?

This phenomenon is a function of the carbon dioxide present in the RO feed water. Because carbon dioxide is a gas, it is not rejected by an RO membrane. Hence, the permeate will contain carbon dioxide if the

feed water contains it. However, the membrane rejects carbonate and bicarbonate which are in equilibrium with carbondioxide in the feed water. Because carbonate and bicarbonate are rejected by the membrane, a new equilibrium occurs in the RO permeate, reducing the pH (see Chapter 9.8 and equation 9.1).

17.2 Operational

17.2.1 When is it Time to Clean an RO Membrane?

Cleaning an RO membrane is generally based on the normalized permeate flow or the pressure drop (see Chapter 13.2.1). When the normalized permeate flow drops by 10%–15% or the pressure drop increases by 10%–15% from start up, it is time to clean. Waiting too long to clean membranes will most certainly result in some permanent fouling. Cleaning too often (when it's not yet time to clean) will result in shorter membrane life, due to the destruction of the membrane by the cleaning conditions and chemicals. Thus, cleaning based on a calendar schedule rather than based on performance is not recommended, because inevitably, the membranes will either be cleaned too often or not often enough.

17.2.2 How Long does it Take to Clean an RO System?

A typical two-stage RO skids can take 8–12 hours to clean, depending on the time it takes to heat up the cleaning chemical solutions. If an extended soak time is required, it can take even longer, up to 24 hours, including the soak period (see Chapter 13.2.2). Each stage in a skid should be cleaned independently of the other(s) so as not to contaminate one stage with foulants or scale from another, which is why it may take a day or so to clear an entire system.

17.2.3 What Temperature Cleaning Solution Should Be Used to Clean Membranes?

Membranes should be cleaned at as high a temperature and at pH extremes as recommended by the manufacturer (see Chapter 13.2.2). Studies have indicated that cleaning under these conditions removes more scale and foulants than cleaning at ambient temperature and neutral pH (see Chapters 13.2.3.1 and 13.2.3.3). However, cleaning outside the recommended temperature and pH parameters leads to

membrane degradation, will void the membrane warranty, and should not be attempted without prior approval from the manufacturer.

17.2.4 Can Extended Soak Time Compensate for Cleaning at Lower Temperature, for Example, When the Heater is Not Working?

Cleaning at lower temperatures is not recommended. Cleaning solutions are typically not effective at ambient temperatures and some components of the solution may even precipitate at these temperatures. Hence, a longer soak will not compensate for cleaning at lower temperatures. In fact, during extended soak periods at higher temperatures, a slow recirculation is recommended to maintain temperature.

17.2.5 Should the Low or High pH Cleaning Be Conducted First?

It is strongly recommended that the high pH cleaning be conducted first. Acid cleaners will generally react with silica, organics, and bio-film on the membranes, making these fouling and scaling problems worse. Acid cleaning should only be used first if it is known that calcium carbonate or iron oxides are present on the membrane.

17.2.6 What Should Be Done if Cleaning Does Not Return Performance to Baseline?

It is acceptable if performance returns to within a couple of percent of baseline. However, if performance falls short of baseline by more than 2 – 3%, additional cleaning is recommended. Refer to Table 13.3 for specific foulants and scale removal cleaning solutions. Should these additional cleaning solutions fail to return performance, off-site cleaning is suggested (see Chapter 15.3). Vendors of off-site cleaning services have access to several cleaning solutions and can try different variations to find one that is effective. If off-site cleaning is ineffective, the fouling or scaling problem is probably irreversible. Improvements to the pretreatment system as well as cleaning when normalized data indicate it is time to do so, will reduce the potential for irreversible fouling or scaling in the future.

17.2.7 If the Clean-in-Place Pump cannot Provide the Required Flow Rate, Can the Pump be Run at Higher Pressure to Compensate?

The CIP pump should provide up to 45 gpm at less than 50 – 60 psi per pressure vessel being cleaned (see Chapter 13.2.4.2). If the pump cannot supply this flow rate, operating at higher pressure will not help; it will make the fouling or scaling situation worse. This is because the higher pressure will force particles and cleaning solution irreversibly into the membrane. Cleaning should be conducted at pressures less than 60 psig. A good indicator of pressure is that little or no permeate should be generated during cleaning.

17.2.8 What Should Be Done With Permeate That is Generated During Membrane Cleaning?

The RO membranes should be cleaned at pressures low enough to prevent generation of permeate during the cleaning process. If permeate is generated, the risk is that the high pressure is forcing cleaning solution and foulants or scale into the irreversibly into the membrane (see Question 17.2.7). If any permeate is generated, reduce the pressure and send the permeate back to the CIP tank.

17.2.9 Why is the Permeate Conductivity High after Cleaning the Membranes?

Permeate conductivity is typically higher the nominal after a high pH cleaning of membranes. High pH used during cleaning “loosen” the membrane polymer, making it more permeable to dissolved solids. This is a temporary condition; conductivity should return to nominal within a few hours to a few days. The converse is true for low pH cleaning; the permeate conductivity will be lower than nominal after low pH cleaning.

17.2.10 Why is Chlorine Both Added and then Removed Prior to the RO?

Chlorine (or other disinfectant) is required to minimize the potential for fouling the membranes with microbes (see Chapters 8.2.1 and 8.2.2). Once membranes are fouled with microbes, it is very difficult

to remove them. A free chlorine residual of about 0.5 to 1.0 ppm in the pretreatment system is desirable. Feed water to the RO must be dechlorinated prior to the membranes because the membranes are sensitive to oxidizers, which will degrade the membrane (see Chapters 8.1.4 and 8.2.4). Sodium bisulfite is the preferred method to dechlorinate unless the RO feed water has a high organic concentration, in which case, carbon filtration at a flow rate of 2 gpm/ft³ is recommended. Sodium metabisulfite is typically about 33% active, and the stoichiometric dosage of sodium metabisulfite is about 1.8 ppm per ppm free chlorine. So, the stoichiometric dosage of 33% active sodium metabisulfite is 5.4 ppm. For safety, a factor of 1.5 is used to increase the dosage of sodium metabisulfite to ensure complete elimination of free chlorine.

17.2.11 What Chemicals Can be Used to Disinfect RO Membranes Directly?

For biocide treatment directly on the membranes, DBNPA is a good non-oxidizing biocide (see Chapter 8.2.5.2). For clean membranes, a dosage of about 100 ppm for 30 minutes 2 to 3 times per week is recommended. For heavier fouling, 100 ppm for 60 minutes 2 to 3 times per week should be fed. Alternatively, DBNPA can be fed continuously at about 2–3 ppm. Note that once biofouling gets out of control, it will be very difficult for DBNPA to work, because it is a non-oxidizer and cannot penetrate biofilm. DBNPA works best as a preventative treatment.

Isothiazolone can be used as a cleaner, but not for slug treatment, as it requires a longer contact time than DBNPA (see Chapter 8.2.5.3). When cleaning with isothiazolone, it should be allowed to contact the membrane for at least 4 hours.

17.2.12 Why does the RO Trip Off on Low Suction Pressure?

Low suction pressure is typically a result of inadequate water supply to the RO feed pump caused by upstream demand starving the RO system. Upstream demands include filter backwash water and water diverted for other applications within the facility. Starving of the RO due to equipment backwashing upstream is a system design flaw. Diversion of feed water usually occurs during installation or

even after, as the need for fresh make-up water grows for other applications within the facility. These issues should be considered during the design phase and/or prior to installation of the RO system.

17.2.13 Should RO Feed Water be Heated?

There are advantages and concerns with heating RO feed water. Heating the water, particularly in the winter time when surface waters tend to be cooler than in the summer months, will reduce the energy required to pressurize the water (see Chapter 9.2). On the other hand, heating water will encourage microbial growth and perhaps microbiological fouling of the RO membranes. These two conditions should be carefully evaluated before heating water is considered. A variable frequency drive or VFD is recommended for applications where there is a significant temperature difference from summer to winter (see Chapters 6.2 and 9.2).

17.2.14 What Limits Recovery by an RO?

The recovery of feed water as product water is a function of several factors. These factors include the concentration of scale formers in the feed water and the design of the RO array (see Chapters 3.7 and 5.1).

- Scale formers limit recovery to the saturation concentration of the scale formers involved. Projections using ion product, solubility products, and LSI should be prepared prior to completing design of the RO system to determine the likelihood for scaling (see Chapter 3.7). Antiscalants can be used to delay or even eliminate scaling so that higher recoveries can be achieved (see Chapter 8.2.3).
- System design plays a role in determining acceptable recovery by an RO. Flow rates per pressure vessel, recovery per module, and BETA values must all be taken into account when considering acceptable recovery by the RO system (see Chapters 9.4, 9.5, and 9.6). The higher the recovery of the RO system, the closer concentrate flow rates and individual module recoveries come to reaching limits recommended by membrane manufacturers.

17.2.15 How do I Start Up an RO?

Starting up an RO, particularly when new membranes have been installed, needs to be done carefully to prevent water hammer from crushing the membrane modules (see Chapter 6.2) First, before anything is started, the start-up procedures provided by the equipment vendor should be read completely and understood. Then, to prevent water-hammer damage to the membranes, the concentrate and permeate valves should be wide open at start-up. *Never start the RO system with the concentrate valve closed and then opening it until the desired recovery is reached.* The RO feed pump should be started slowly, increasing the pressure at a rate no greater than 10 psi per second. If a variable frequency drive is used, it can be adjusted to start up slowly. If not VFD is installed, and a centrifugal pump is being used, the concentrate valve should start open and then be closed down slowly until the desired recovery and feed pressure is reached, making sure that the pressure increases at an acceptable rate. The use of older, positive displacement pumps requires a pulsation dampener and a slow start, using the concentrate and pump recycle valves (both of which are to start wide open) to adjust the recovery and feed pressure.

To prevent damage to the membranes, they should be properly shimmed and the thrust ring correctly installed (see Chapter 4.3.3 and Question 17.3.4). The shims and thrust ring will minimize or prevent movement of the modules during start-up and shut-down of the RO system.

17.2.16 Do RO Membranes Need to be Preserved When Taken Off Line?

First, when membranes come off line, they should be flushed with either permeate water or low-pressure feed water (see Chapter 13.1.1). This will reduce the concentration of ions and any suspended solids on the feed side of the membrane, thereby minimizing the potential for fouling or scaling the membrane while idle. The next step(s) depends on how long the membranes will be off line.

- **Short-term idling:** Short-term idling includes membranes that are off line for no more than 48 hours. An automatic flush event should occur at least once every 24 hours. No other steps need to be taken to preserve the membranes.

- Short-term storage: Short-term *in situ* storage includes membranes that are off line for longer than 48 hours but less than 2 weeks (see Chapter 13.1.1).
 - Membranes should be flushed as described above. All air should be vented from the system while flushing. This can be accomplished by overflowing the flushing solution through the utmost point of the highest pressure vessel.
 - Once the vessels are filled, all valves should be closed to prevent air from entering the membranes. These steps should be repeated every 5 days.(2)
- Long-term storage: Long-term *in situ* storage includes membranes that are off line for longer than two weeks (see Chapter 13.3.2). It is necessary to clean the membranes prior to storage. Standard cleaning techniques should be applied (see Chapter 13.2). After cleaning, the following preservations procedures should be conducted:
 - Circulate a 1 to 1.5% solution of sodium metabisulfite through the membranes, completely filling the pressure vessels. To ensure that the vessel are completely filled, the solution should be allowed to overflow through an opening located at the utmost point of the highest pressure vessel being filled.
 - Once the pressure vessels are filled with the bisulfite solution, all valves should be closed to prevent the oxygen in the ambient air from oxidizing the sodium metabisulfite.
 - The pH of the preservative solution should be checked once a week. The solution should be changed out when the pH reaches 3 or lower.
 - The preservative solution should be changed at least once per month during colder weather (less than 80°F), regardless of its pH. During warmer weather, the solution should be changed every two weeks.

During long-term *in situ* storage, the following precautions should be taken:

- Membranes should not be allowed to dry out. Dry membranes irreversibly lose flux.

- Temperature extremes should be avoided. The system must be kept free of frost (typically greater than 5°C) and should not be allowed to exceed 45°C (consult with membrane manufacturer for temperature ranges for specific membranes). Cooler temperatures are preferred, as lower temperatures minimize microbial growth.

17.2.17 Is there a Shelf Life for Reverse Osmosis Membranes?

Shelf life depends on the condition of the membrane when they were stored. New membranes that are stored in their original, unopened bags have a shelf life of about 1 year.^{3,4} Membrane warranties typically start at system start-up or 1 year after shipment, whichever comes first.³ Membranes that were wet tested prior to shipment from the factory should be inspected every three months for biological growth and for pH. If biological activity is found (the preservative solution is not clear), after six months of storage, or if the pH of the preservative solution drops below 3, the modules should be stored in fresh preservative solution, as described below. Membranes that were not wet tested ("dry" membranes) should also be inspected regularly for biological growth. If growth is found, they should be soaked in a preservative solution as described below.

Membranes that have been used can be removed from the RO skid and stored. They should be cleaned prior to storage. Once removed from the pressure vessel, each membrane module should be soaked in a 1% solution of non-cobalt activated sodium metabisulfite solution mixed with deionized water such as RO permeate. For greater disinfection and protection, Toray recommends soaking on a 0.2 to 0.3% solution of formaldehyde for membranes that have been in operation for more than 72 hours.⁵ (Note that storage in formaldehyde may result in permanent flux loss. Formaldehyde is a poison to humans, and, therefore, is not recommended for food-related applications. Consult with the membrane manufacturer before using preservatives other than sodium metabisulfite.) The modules should be soaked in the vertical position for one hour. After soaking, allow the module to drip and then storage it in an oxygen-barrier plastic bag. There is no need to fill the bag with the preservative solution, as the

moisture in the module is adequate. Modules can be stored for six months using this method. Stored membranes should be inspected every three months for biological growth and for pH. If biological activity is found, after six months of storage, or if the pH of the preservative solution drops below 3, the modules should be soaked in fresh preservative solution and sealed in a new oxygen-barrier bag. Membrane should be cleaned in a high-pH solution when it comes time to return them to service.

Membrane modules should always be stored in a cool, dark place out of direct sunlight and kept from freezing. Wet-tested membranes should be stored at no lower than about 5°C to prevent freezing of the sodium metabisulfite preservative solution (FilmTec membranes can go to -4°C).⁶ Dry membranes will not be affected by freezing temperatures. (Note that once wetted, membranes should not be allowed to dry out, as irreversible loss of flux may occur.)

17.2.18 What is the Difference Between Membranes that Have Been Wet Tested and those that are Dry?

Brackish water membranes can be shipped from the manufacturer wet or dry. Wet membranes have been performance tested at the factory. However, testing is usually conducted for shorter (hours) versus longer (days) periods. As noted in Figure 14.2 (Chapter 14.3.2), there is a period of time after start up during which membrane performance is not stable due to compaction. Flux and rejection both decrease during this period. Unless a membrane is wet tested until stable performance is achieved, the performance specifications for that membranes based on the wet test are not accurate.⁷

Dry membranes may have been leak tested with air or they may not have been tested at all.

17.2.19 What is the Impact on the RO If the Pretreatment System Fails, for Example, If the Softener Leaks Hardness?

Any failure in the pretreatment system will be registered on the RO membranes as fouling or scaling or degradation (see Chapter 12). Failures include:

- Malfunction of chemical pretreatments, including chlorine, sodium bisulfite, and antiscalant leading to biofouling, degradation, and scaling, respectively
- Channeling through multimedia filters leading to fouling of the membranes
- Hardness leaking from softeners leading to scaling of the membranes
- Particle unloading from cartridge filters due to high pressure drop leading to fouling of the membranes

Fouling will usually affect the first membranes in the first stage of the RO system. Pressure drop will increase across this stage and operating pressure will increase. Normalized permeate flow will decrease over the first stage. Degradation will also primarily affect the first stage. Salt passage will increase and operating pressure will decrease. Normalized permeate flow will increase over the first stage. Scaling will affect the last stage of the RO system. Salt passage and operating pressure will both increase. Normalized permeate flow will decrease over the last stage. Refer to Chapters 3.6, 3.7, and 12 for additional information.

To prevent these failures, constant monitoring of the pretreatment system is necessary. Alarms should be installed on critical systems, such as the ORP associated with the sodium bisulfite feed. Particle monitors could be used to detect channeling or carry over through filters. Hardness analyzers with alarm should be installed on the effluent from softeners.

17.2.20 Can Different Types of Membranes Be Used in an RO Unit?

Different membranes can be mixed in a single RO unit, but is usually not recommended. One case where membranes are mixed is in low-pressure systems. In a low-pressure system, the water flux can drop off significantly through the last few membrane modules as the osmotic pressure of the feed approaches the difference between the applied pressure and the pressure drop (driving force) in the pressure vessel. This situation is common in low-pressure municipal applications where many systems have 7 membrane modules in series in a pressure vessel.⁸ In this case, the last 2 or 3 membranes in the last stage can be replaced with low-energy (high flow) membranes (see Chapter 4.4.2.1). These low-pressure membranes usually sacrifice rejection, but high rejection is not as critical for municipal

applications where 80 – 90 ppm TDS product water is acceptable. Note that low-pressure membranes should never precede standard-pressure membranes in an RO system. This is because water will follow the path of least resistance and a disproportional amount of water will flow through the lead membranes, in effect “starving” the later, standard-pressure membranes.

The other case where membranes are sometimes mixed is during emergency situations where some but not all membranes have been damaged in some way and need replacing. In this case, it is not uncommon for membranes from different manufacturers to be combined in a single RO unit. If this is a necessity, the interconnectors should be compatible with each membrane (see Chapter 4.3.3). Because of the variations among interconnections, membranes from different manufacturers are generally not combined in the same stage of an RO system. Also, performance specifications should be similar for all the different membranes used in an RO system.

17.3 Equipment

17.3.1 What is the Footprint For an RO System?

While the footprint of an RO system will obviously vary with size of the system, there are some generalities that can be made. The length of the RO system depends on how many membrane modules are in series in the pressure vessels. Table 17.1 lists the approximate size of an RO skid as a function of the number of modules in the pressure vessels. Note that sizes may vary depending on the manufacturer.

Additionally, there are “work zone” areas that should be allowed for in the plant layout of the RO skid. At each end of the RO skid, a zone of a minimum of 4 feet (6 feet is best) should be allotted for loading and unloading membrane modules. The front side of the skid should have about 4 feet allotted for pump maintenance and access to the controls and instruments. The backside of the skid requires no access and the skids can be placed 1 to 2 feet away from adjacent equipment.

17.3.2 What is a Variable Frequency Drive Used For?

A variable frequency drive (VFD) adjusts the speed of a motor to alter the discharge pressure (see Chapter 6.2). This is useful when

Table 17.1 Approximate RO skid size as a function of number of modules per pressure vessel.

Number of Modules in Series	Length, inches	Width, inches	Height, inches	Array	Capacity, gpm
3	168	32	77 – 89	3:2:1	30 – 100
4	226	57	80 – 91	4:3:2 3:2:1 6:4:2	100 – 200
6	280	66	115	8:4 10:5	320 – 400
6	280	94	128	14:7 18:9	560 – 720

there are significant variations in feed water temperature with changes of seasons. Because an RO membrane requires lower pressure at higher temperature to force water through the membrane, energy can be saved during the summer months if a VFD is employed (see Chapter 9.2). Thus, during the warmer months, the VFD can be used to “dial back” the discharge pressure, reducing the energy required to operate the system.

17.3.3 What is the Difference Between Pleated, String-Wound, and Melt-Blown Cartridge Filters?

All three types of cartridge filters are acceptable for pretreatment to RO membranes (see Chapter 6.1). Pleated cartridge filters are typically used in higher-purity applications such as pharmaceuticals and microelectronics. String-wound filters are just as they sound; material such as polypropylene in string form is wound around a central core. These filters suffer from particle unloading at higher pressure drops and require a slower velocity through them than other types of cartridge filters, typically 2–3 gpm per 10-inch equivalent rather than 5 gpm per 10-inch equivalent for pleated and melt-blown filters. The melt-blown variety is thermally bonded polypropylene microfibers and is typically denser near the core than at the outside.

This allows for particles to be trapped throughout the cross section of the filter similar to a depth filter.

17.3.4 What is the Correct Way to Install Shims and the Thrust Ring?

Shims and the thrust ring are used to protect the membrane modules from moving around in the pressure vessel during start-up and shut-down of the RO system (see Chapter 4.3.3).

The thrust ring is installed at the concentrate end of the pressure vessel before any membranes are installed. Consult with the manufacturer of the pressure vessel for correct orientation of the ring.

Shims are installed at the feed end of the module/pressure vessel assembly (refer to figures 4.18 and 4.19). Because pressure vessels are constructed with slight variations in length (known as “freeboard”), membrane modules can slide during pressurization and depressurization. Shims are installed between the face of the lead module and the adapter hub to prevent this motion. Membrane modules should be pushed completely against the thrust ring prior to installation of the shims. Shims are washer-like plastic rings that may be purchased from the pressure vessel manufacturer or fashioned out of PVC (must be free of burrs and be cut parallel to work properly).

17.3.5 How Should the Cleaning Pump be Sized?

The cleaning pump should be sized to handle 45 gpm per 8-inch pressure vessel to be cleaned at a pressure less than a 50–60 psig (see Chapter 13.2.4.2). For example, given an 8:4-6M array, the pump would have to provide cleaning solution for a maximum of 8 pressure vessel at one time. At 45 gpm per pressure vessel, the pump needs a discharge flow of 360 gpm.

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Unit Equivalents and Conversions

Parameter	Value	Equivalent Values
Mass	1 lb _m	= 453.593 g = 0.453593 kg
Length	1 ft	= 0.3048 m = 30.38 cm
Volume	1 ft ³	= 7.4805 gal = 28.317 l = 0.028317 m ³
Pressure	1 psig	= 0.06895 bar = 6895 Pa
Power	1 hp	= 754 watts
Flux	1 gfd	= 0.58865 l/m ² hr = 0.0005887 m/hr
Temperature	1°F	= °C*1.8 + 32

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